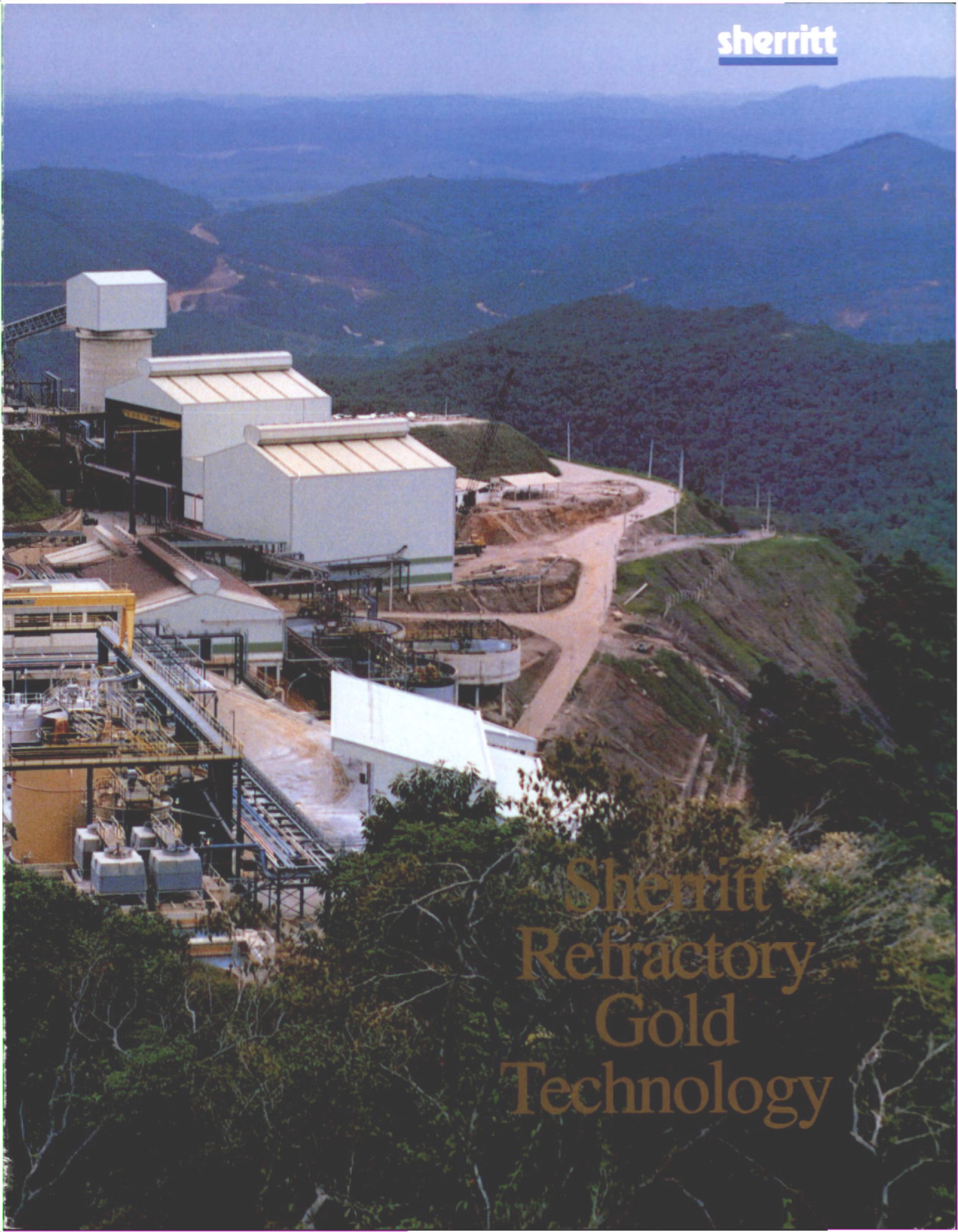


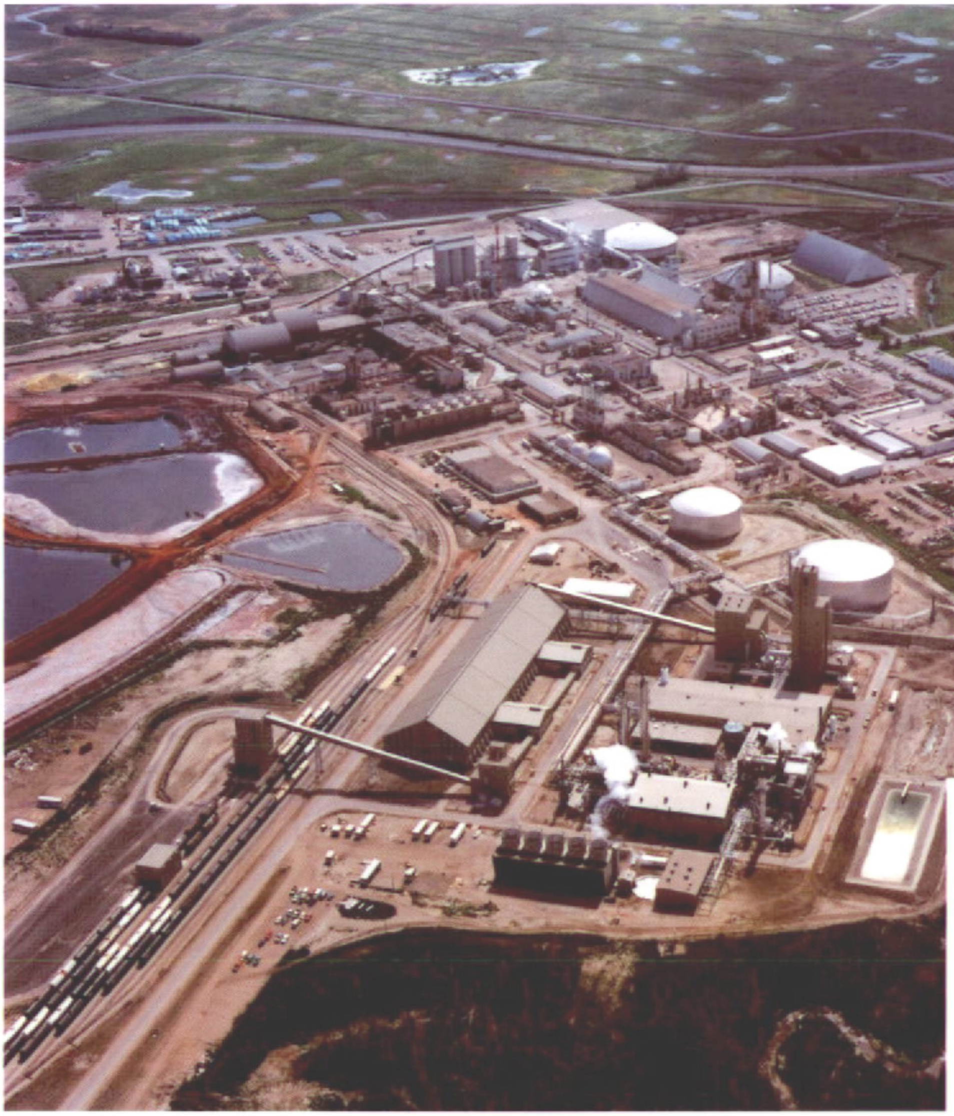
sherritt

Sherritt Refractory Gold Technology

About Sherritt

Sherritt Gordon Mines Limited operates an integrated metallurgical-chemical complex at Fort Saskatchewan, Alberta, where it manufactures nitrogen and phosphate fertilizers and sulphur chemicals, refines nickel and cobalt, and produces various metal products including composite powders and coinage materials. Sherritt also operates two mines in northern Manitoba: Ruttan, which produces copper and zinc concentrates, and MacLellan, which produces gold and silver bullion.

Sherritt's External Technology Group actively licenses the company's proprietary pressure hydrometallurgical and sulphur chemical processes and product technology worldwide. External Technology also provides contract research and specialized engineering services in support of these processes and supports the development of new process and product technology.

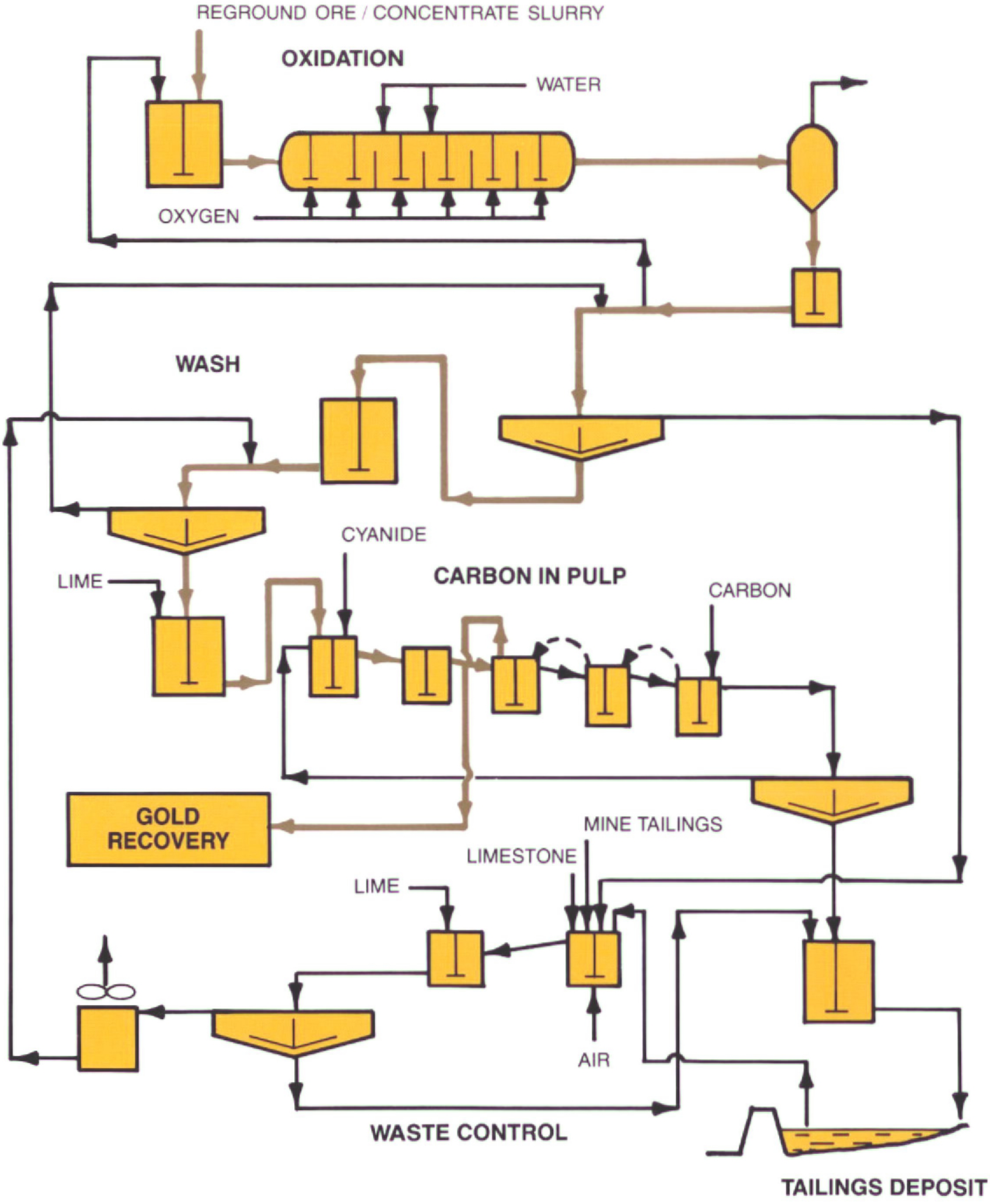


Sherritt's Fort Saskatchewan Complex

MacLellan Mine surface facilities. Gold ore from the mine is processed to produce doré bullion at the nearby mill in Lynn Lake, Manitoba.

Cover Photograph: Sao Bento Mineracao S.A. gold plant in Brazil.
Sherritt provided process development, engineering, and commissioning services for this facility.

FLOWSHEET FOR PRESSURE OXIDATION OF REFRACTORY GOLD ORES


REGROUND ORE / CONCENTRATE SLURRY
OXIDATION
WATER
OXYGEN
WASH
CYANIDE
CARBON IN PULP
CARBON
LIME
GOLD RECOVERY
MINE TAILINGS
LIMESTONE
LIME
AIR
WASTE CONTROL
TAILINGS DEPOSIT

THE SAO BENTO GOLD PROJECT

PRESSURE OXIDATION PROCESS DEVELOPMENT

by

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This paper was presented at Projects '88, 18th Annual Meeting, Hydrometallurgy Section of the Metallurgical Society of the Canadian Institute of Mining and Metallurgy Edmonton, Alberta May 1988

1988-253

THE SAO BENTO GOLD PROJECT - PRESSURE OXIDATION PROCESS DEVELOPMENT

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ABSTRACT

Preconcentration of Sao Bento ore results in high recovery of gold in a flotation concentrate. The bulk of the gold is contained in auriferous arsenopyrite, pyrrhotite, and pyrite and is not recoverable by conventional cyanidation. Acid pressure oxidation proved to be an effective method for liberating the refractory gold. Batch and continuous pressure oxidation testwork conducted by Sherritt are described, including the problems and solution developed for coping with elemental sulphur formation and sulphide agglomeration. Miniplant treatment of the concentrate in a continuous integrated circuit comprising pressure oxidation; residue washing; gold recovery by cyanide leaching and carbon in pulp adsorption; and, waste treatment is described. A brief description also is provided of Sao Bento Mineracao's commercial plant in Minas Gerais, Brazil, which commenced operations in late 1986.

Permission is hereby granted to publish, with appropriate acknowledgements, excerpts or summaries not to exceed one-fourth of the text of the paper.

Presented at Projects '88, 18th Annual Meeting,
Hydrometallurgy Section of the Metallurgical Society of the
Canadian Institute of Mining and Metallurgy
Edmonton, Alberta, May 1988

INTRODUCTION

The Sao Bento gold deposit is located in the Minas Gerais province of Brazil, approximately 500 km north of Rio de Janeiro. The major sulphidic minerals in the ore include arsenopyrite, pyrite, and pyrrhotite. A significant portion of the gold is refractory to recovery by conventional cyanidation because of its intimate association with the arsenopyrite and pyrite. General Mining Union Corporation Limited (Gencor) had evaluated several alternatives as a pretreatment to improve the extractability of the gold.

Batch pressure oxidation tests were conducted on samples of ore and flotation concentrate and a brief continuous pressure oxidation run was conducted on the concentrate at the Sherritt Research Centre in May 1983 on behalf of Gencor. The tests confirmed the amenability of the feedstocks to pressure oxidation, as the gold extraction was improved from about 30% attainable by direct cyanidation, to the range of 96 to 98% from both the ore and concentrate after pressure oxidation. Economic feasibility studies of the various pretreatment alternatives indicated that, although the capital and operating costs for a concentrate roasting facility were somewhat lower than for a pressure oxidation facility, the differential was more than offset by the higher gold recovery offered by the latter alternative.

The next phase of the overall development program for a pressure oxidation process was conducted on flotation concentrate. The scope of work included unit operations testing of the various processing steps, followed by a 250 h continuous fully integrated run in the miniplant including feed preparation, pressure oxidation, residue washing, gold leaching and recovery, waste control, and tailings deposition. The objectives of the laboratory program were to demonstrate the process and maintain high gold recoveries and acceptable effluent quality. A further objective was to design and execute the work such that sufficient data would be generated and interpretation made, which together with Sherritt's engineering and commercial operating experience on autoclave circuits would allow the project team to proceed with the design of a commercial plant without the need for further piloting on a larger scale or for more extensive miniplant operations. The test program was initiated in November 1983, and was completed by mid February 1984.

This paper describes the batch testwork, the unit operations testing, including the problems associated with the formation of elemental sulphur and its agglomeration of unreacted sulphides during pressure oxidation, the resolution of the above problem, and the miniplant run which confirmed the integrity of the flowsheet and established the technical viability of the entire process.

BATCH TESTING

Batch pressure oxidation amenability tests were conducted on samples of flotation feed and flotation concentrate, whose chemical compositions are given in Table I.

Table I. Chemical Composition of Batch Test Feed Materials

Material	g/t Au	g/t Ag	% As	% Ca	% Fe	% Mg	% S	% Zn
Flotation Feed	10.1	0.7	3.6	2.1	20.6	1.6	6.4	0.03
Flotation Concentrate	51.0	2.6	19.2	0.8	38.6	0.4	28.0	0.04

The approximate sulphide mineralogy of the concentrate was as follows: 42% (by weight) arsenopyrite, 21% pyrite, and 21% pyrrhotite. In addition to the carbonates of calcium and magnesium, the flotation feed contained an appreciable amount of siderite, much of which was rejected to the tailings in the flotation. Gold extractions by direct cyanidation of the flotation feed and concentrate, reground to 88% passing 44 μm, were 43 and 31%, respectively.

Batch pressure oxidation tests were conducted to establish the effects of oxidation temperature, retention time, oxidation additives and feed regrind on the liberation of refractory gold and to define criteria for continuous testing. As with most sulphidic feeds evaluated by Sherritt, the release of the refractory gold was directly related to the degree of sulphur oxidation (1,2). This relationship is shown in Figure 1. Sulphur oxidation kinetics were enhanced by a feed regrind and an increase in temperature (200°C vs 185°C). The higher temperature provided marginally higher gold extraction at comparable levels of sulphur oxidation.

The pressure oxidation pretreatment provided equally effective liberation of the gold from both the flotation feed and the concentrate. Also, the flotation feed contains sufficient sulphur to enable autogenous oxidation and to meet the apparent acid demand of the acid consuming gangue components. These factors would normally favour direct pressure oxidation of the flotation feed (3). The processing of flotation concentrates, however, was selected for the Sao Bento feeds for two major reasons. Firstly, recovery of gold to the concentrate in flotation was high, such that losses to the tailings were minor. Secondly, the bulk of the carbonate content comprised siderite, which was not readily decomposed in an atmospheric acidulation of the feed, but which did react under autoclave conditions. Losses of oxygen in the vent gases used to purge carbon dioxide from the autoclave, therefore, favoured processing of flotation concentrate.

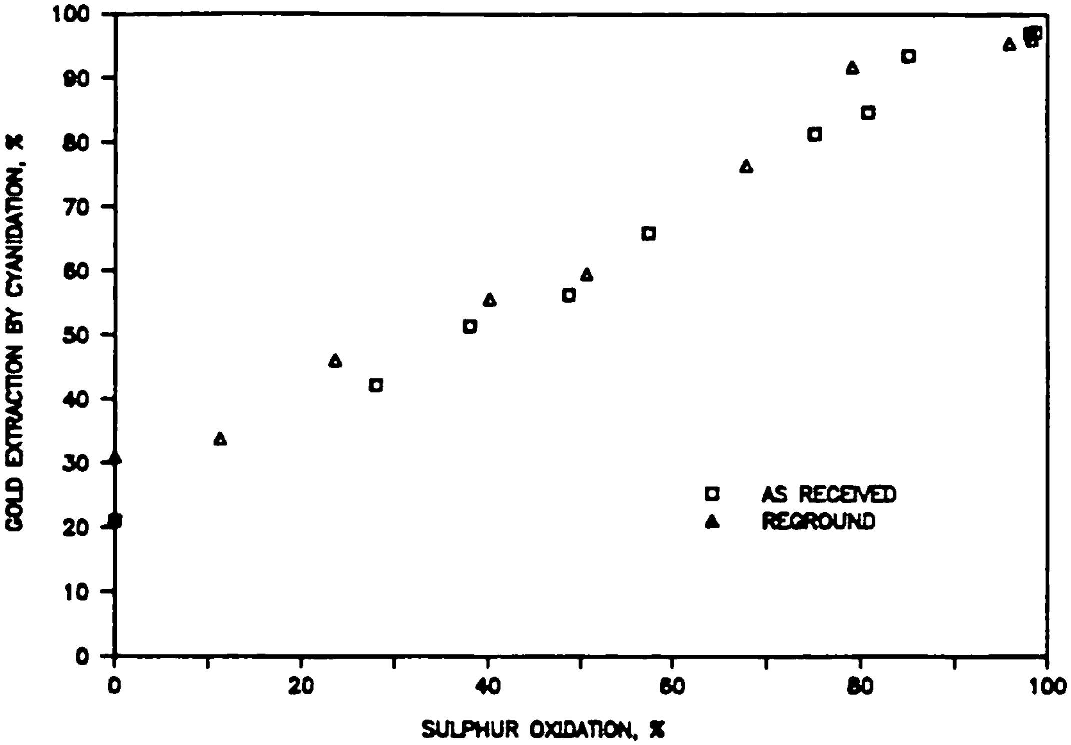

Figure 1. Gold Extractability as a Function of Sulphur Oxidation at 185°C

A brief continuous pressure oxidation run on the concentrate was also conducted as part of the preliminary study. The run, of 13.5 h duration, was conducted in a 10 L four cell continuous autoclave, at a nominal solids content of 14%, at 185°C, with a solids retention time of 2 h. Gold extraction from the oxidized solids, at 94.5 to 95.0%, was slightly lower than in the batch tests. This was attributed to incomplete sulphur oxidation due, in part, to flotation and short circuiting of sulphides, and also to the formation of elemental sulphur. At shutdown, a small quantity of plus 125 μm sized "minipellets" was recovered from the first cell. The solids, which contained 130 g/t Au, 21% As, 32% Fe, 41% S, and

10.7% S°, were upgraded in gold content and contained virtually no pyrrhotite. This suggested that the elemental sulphur had originated primarily from pyrrhotite.

Formation of small agglomerates had occurred in a previous run on a pyrrhotite and base metal sulphide containing arsenopyrite concentrate but had largely been controlled or minimized through the use of oxidation additives.

CONTINUOUS UNIT OPERATIONS TESTING

Unit operations testing included continuous pressure oxidation, two stages of residue washing in a countercurrent decantation circuit, continuous gold recovery by cyanidation and carbon in pulp adsorption, and precipitation of metals and associated sulphur from the acidic waste solution from the pressure oxidation in a neutralization circuit. Data derived from the unit operations testing were used to define the criteria for the integrated run.

The concentrates used in the continuous integrated runs were lower in grade than the one employed in the initial batch tests, because of 'pulling harder' during flotation to maximize gold recoveries. The concentrates contained in the range of 29.1 to 35.4 g/t Au, 11.6 to 14.8% As, and 18.6 to 23.4% S. Batch pressure oxidation amenability testing confirmed a response similar to that of the high grade concentrate.

Continuous Pressure Oxidation

Six continuous oxidation runs were conducted in the multicompartment horizontal autoclave. The autoclave, described in previous publications (1,2,4,5,6,7), had a working volume of about 26 L. The oxidation runs ranged from 10 to 57 h and varied in scope, from solely continuous oxidations with no recycle streams, to oxidation runs in conjunction with the CCD wash circuit, and to fully integrated runs incorporating the CCD wash, gold recovery, and waste treatment circuits prior to the major integrated run.

The first two oxidation runs were conducted in the temperature range 180 to 185°C, but were terminated after 24 and 44 h, respectively, because of a heavy buildup of agglomerated solids in the first and second compartments. The agglomerates contained about 95 g/t Au, 26% As, 31% Fe, 40% S and 16% S°, and comprised predominantly unreacted arsenopyrite and pyrite, agglomerated by elemental sulphur. Neither a concentrate regrind, increased agitation, nor increased levels of additives were effective in preventing the agglomeration.

At this stage, the testwork on the Sao Bento concentrate was transferred into Sherritt's internal research program to resolve the problem of sulphide agglomeration before resuming the Gencor program. Several qualitative batch oxidation tests were conducted to test a number of alternatives. The one that showed the most promise was the recycle of oxidized solids, at a recycle solids:concentrate weight ratio in the range of 3:1 to 5:1, to maintain a high pulp density to promote sulphur suspension and dispersion (1,6,8).

This concept was further tested and its effectiveness was confirmed in a 57 h continuous run. The feed to the autoclave was a 50% solids slurry consisting of a blend of recycled oxidized solids and fresh concentrate in a weight ratio of 3.5:1. The run consisted of five operating periods during which the effects of temperature, concentrate regrind, and retention time were evaluated. The results for the period on reground concentrate (96% minus 44 μm), a nominal retention time of 2.4 h, and a temperature of 190°C in the first compartment, decreasing to 185°C in the final compartment are summarized in Table II.

Table II. Continuous Pressure Oxidation of Sao Bento Concentrate

Autoclave Compartment	1	2	3	4	5
Sulphur Oxidation, %	64.4	84.6	91.1	93.1	98.1
Cyanidation Residue, g/t Au	2.58	1.45	1.10	0.92	0.94

The use of the recycle of oxidized solids was effective in preventing the problem of agglomeration and enabled trouble free continuous operation. The median gold extractability from the oxidized solids during the above period was 96.6%.

In view of the high recycle of solids necessary to effect successful continuous oxidation of the concentrate, the possibility of direct pressure oxidation of the ore was reviewed. Preference was given to the oxidation of the concentrate for the following reasons: preparation of flotation concentrates ensured a more consistent composition for the pressure oxidation; the processing of concentrate would require a smaller regrind circuit; the treatment of concentrate would result in a smaller gold recovery circuit and lower cyanide consumption; and, the direct oxidation of ore would require a large second wash thickener, and probably an additional oxidation thickener to recover acidic oxidation liquor to react with the high carbonate content of the ore prior to autoclaving.

One more attempt was made at continuous oxidation of a concentrate alone, this time one low in pyrrhotite content. Although the 10 h continuous run proceeded without incidence of agglomeration, the metallurgical performance was unsatisfactory and this, combined with the unacceptably high deportment of gold to the magnetic fraction during the preparation of the concentrate made this approach unattractive.

The final two continuous runs, of 37 and 18 h duration, respectively, and both employing a recycle of oxidized solids, constituted the initial and terminal periods of a 181 h fully integrated run on arsenopyrite concentrates. The intervening 126 h was on concentrate from Giant Yellowknife Mines, conducted as part of Sherritt's internal research program, and described in previous publications (1,6). The integrated circuit included feed slurry preparation (incorporating the recycle of oxidized solids from the wash circuit), continuous pressure oxidation, two stages of countercurrent decantation washing, cyanidation and carbon in leach gold recovery, and neutralization of the wash circuit liquors, with recycle of the metal free water to the wash circuit.

The first portion (37 h) was at a lower solids recycle ratio (3:1), lower solids content, and marginally shorter oxidation time than employed in the first continuous run with the recycle. Gold extractability from the autoclave discharge solids was 95.8%. The last portion (18 h) under similar conditions, was designed to purge the previous solids from the oxidation and wash circuits prior to the major integrated run.

The autoclave discharge slurry was washed in the two stage CCD circuit, at a wash ratio in the 16 to 18:1 range. Percol 351 addition levels were 62 and 15 g/t concentrate, respectively, and the underflow slurries contained 47 to 54%, and 64% solids, respectively.

Continuous Gold Recovery Testing

Prior to continuous testing, batch kinetic studies were conducted to examine the effects of temperature, pH, solids content, cyanide concentration, and time, and to establish guidelines for the continuous run. The continuous test was conducted in a circuit consisting of a liming and pulp density adjustment tank, a leach tank, and twelve carbon in leach stages. Slurry retention times were approximately 3 h in each of the liming and leach tanks, and about 35 min in each CIL stage. The feed to the gold recovery circuit was the second wash thickener underflow slurry from the 37 h continuous oxidation run, containing in the range of 29.0 to 33.7 g/t Au.

The slurry was limed to a pH in the 11.4 to 11.6 range and was diluted with water, from 60 to 40% solids. The liming and the leach operations were both at 40°C. The sodium cyanide addition level was maintained at 2.6 kg/t. Gold extraction in the leach was in the 90.8 to 95.3% range, with corresponding residues containing 1.49 to 2.96 g/t Au. Each CIL stage contained 32 g/L carbon. The leaching of the gold continued in the CIL circuit, attaining extractions of 95.0 to 95.3% in the first stage, and 95.2 to 96.4% in the discharge solids from the twelfth stage. The extraction was essentially completed within about seven stages. The leach solution entering the CIL section contained between 13.2 and 17.5 mg/L Au. In the CIL circuit, gold loading in excess of 10 kg/t was readily achieved, while maintaining barren gold tenors at acceptable levels, within a reasonable number of stages.

Waste Management Testing

The weakly acidic waste solutions from the wash circuit contain metals dissolved during the oxidation which must be removed prior to recycle of the water to the process or discharge as plant effluent. These solutions were treated in a neutralization circuit consisting of four tanks in series. Limestone was added to the first tank, to a pH in the range 3.5 to 4.4, and lime was added to the third tank, to pH 10.0 to 11.4, to precipitate the metals as their respective hydroxides, hydrated oxides and arsenates, and to remove the associated sulphate and free acid as gypsum. The neutralization circuit was operated during the 187 h continuous oxidation run on the Sao Bento and the Giant Yellowknife concentrates. The median concentrations of the principal contaminants in the neutralization feed and in the waste waters produced in the neutralization circuit are given in Table III.

Table III. **Neutralization of Acid Waste Water from Continuous Oxidation**

Solution	As	Cr	Co	Cu	Fe	Pb	Hg	Ni	Se	Zn
Feed, mg/L	260	n.a.	15	52	4010	<50	n.a.	61	n.a.	185
Product, mg/L	0.08	0.015	0.007	0.04	0.06	0.1	<0.01	0.02	<0.2	<0.01

Several methods of dewatering the neutralization discharge slurries were tested, including thickening, belt filtration and direct settling in a tailings box. The addition of flotation tailings to the neutralization circuit enabled effective dewatering of the discharge slurry in a thickener alone. The treatment of the solid wastes from the neutralization and the gold recovery circuits was addressed in the integrated run.

INTEGRATED RUN

The feed to the autoclave was a 1:3 blend of reground concentrate and recycled first wash thickener underflow slurry of oxidized solids. The feed to the neutralization circuit included flotation tailings to supply a portion of the neutralization reagent requirements. Initially, liquid-solid separation of the neutralization waste products was done by thickening followed by filtration. Adequate dewatering of the slurry was achieved by thickening alone allowing removal of the belt filter from the circuit. The product water from the neutralization circuit was largely recycled to supply the requirements of the wash circuit and, after 90 h of the run was also used for dilution of the oxidation slurry in the autoclave. In the gold recovery circuit, the limed feed pulp was diluted, initially with water, prior to cyanidation. Advantage was taken of the favorable settling characteristics of the CIL discharge slurry to supply the water requirements for the limed slurry dilution with recycled cyanide containing barren CIL decant water.

Pressure Oxidation

The continuous pressure oxidation circuit was operated for 250 h. The oxidation run consisted of five periods, during which the effects of temperature, slurry pulp density and oxygen stoichiometry were evaluated. Concentrate, reground to 93% minus 44 µm was oxidized. The chemical composition of the concentrate is provided in Table IV.

Table IV. Chemical Composition of Concentrate for Integrated Run

%, g/t Au								
Au	Al	As	Ca	CO_2	Fe	Mg	SiO_2	S
32.3	1.62	12.2	0.86	5.25	34.5	0.96	11.7	20.4

The approximate weight content of the sulphides was 27% arsenopyrite, 26% pyrrhotite, and 11% pyrite, with the three minerals accounting for about 26%, 46%, and 28%, respectively, of the sulphidic sulphur in the concentrate.

The feed to the autoclave was a 55% solids slurry consisting of a 1:3 weight blend of concentrate:recycled oxidized solids, made up of 73% solids concentrate slurry combined with first wash thickener underflow slurry, containing 52% solids.

The oxidation parameters and operating conditions for the integrated run are summarized in Table V.

Table V. Oxidation Conditions and Parameters

Period	1	2	3	4	5
Duration, h	42	48	48	30	81
% Solids, Compartment 1	48.2	48.2	44.1	44.8	48.2
Retention time, h	2.24	2.28	2.02	1.99	2.14
Temperature, °C C1 C2 to C5	190 190	185 190	185 190	190 190	190 190

Water was added to compartments C1, C2, and C3 to dilute the slurry to the desired levels, consistent with the selected temperature profiles. Oxygen was sparged into all the compartments.

A summary of the median sulphur oxidation profiles for each period is provided in Table VI.

Table VI. Sulphur Oxidation Profiles

	Sulphur Oxidation to Sulphate, %				
	C1	C2	C3	C4	C5
Period 1	70.6	85.6	93.8	97.0	99.7
Period 2	65.3	83.3	89.5	95.6	97.4
Period 3	68.0	85.6	89.9	97.1	97.9
Period 4	68.9	81.8	90.8	94.7	99.7
Period 5	66.3	82.8	n.a.	96.8	99.0

The oxidation rates in the five periods of the integrated run were similar, and a consistently high degree of sulphur oxidation was attained in the discharge solids throughout the run.

The median cyanide amenability residue assays for the autoclave profile solids from each period are summarized in Table VII.

Table VII. Cyanide Amenability Leaching of Oxidized Solids

	Gold in Amenability Residues, g/t				
	C1	C2	C3	C4	C5
Period 1	3.84	2.45	2.27	1.94	1.94
Period 2	3.91	2.67	2.11	1.78	1.59
Period 3	3.88	2.86	2.28	2.01	1.88
Period 4	3.86	2.73	2.27	2.27	1.86
Period 5	3.56	2.36	n.a.	1.79	1.44

A comparison of the sulphidic sulphur content of the autoclave discharge solids with the gold content of the residues of cyanide amenability leach tests conducted on these solids is provided in Figure 2.

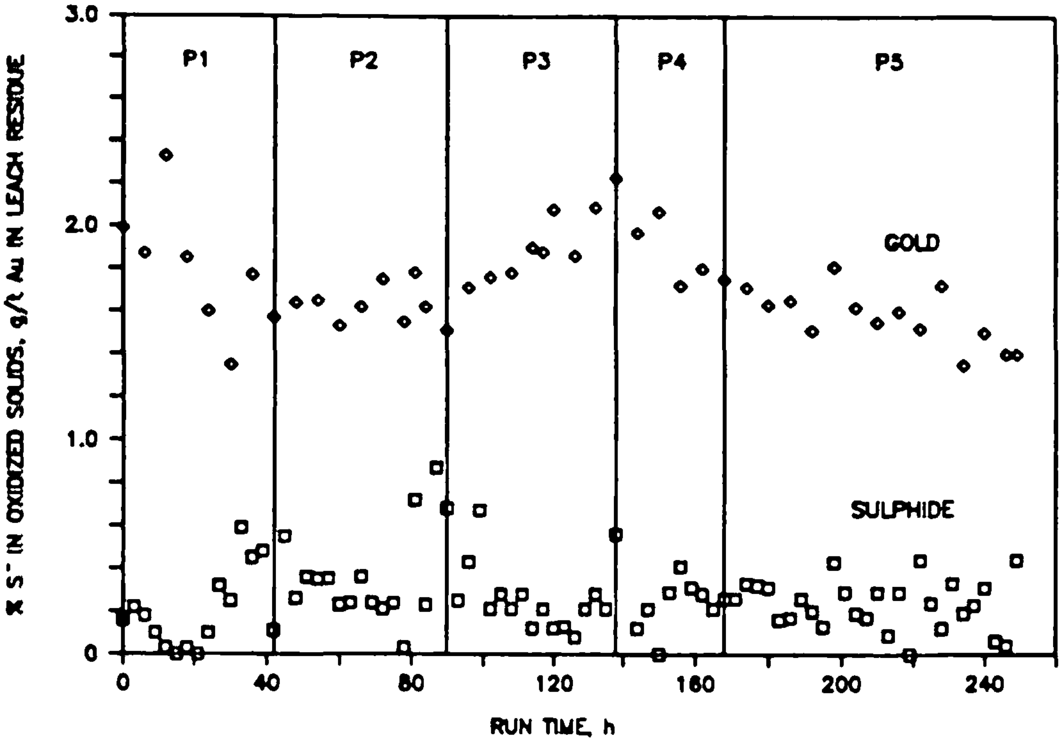

Figure 2. Autoclave Discharge Solids Sulphide and Amenability Gold Profiles

During Period 1, the amenability residues in the initial stages contained high levels of gold, due to residual less responsive Giant Yellowknife solids. The gold extractions improved towards the latter stages of Period 1 and were less variable throughout Period 2. The gold levels in the cyanide amenability residues were in the 1.5 to 1.6 g/t range (95.0 to 95.3% extraction). Period 3 was marked by a gradual deterioration in the gold extraction, decreasing to about 93% (2.23 g/t Au in residue) at the end of Period 3. This was in spite of a high degree of sulphur oxidation throughout the period. The extraction improved gradually during Period 4, after an increase in the C1 temperature to 190°C and continued to improve during Period 5. The major change in the operating conditions in Period 5 had been an increase in the solids content of C1 and of C2 to C5 to 48 to 38%, respectively. The best and most consistent results were obtained in the latter stages of Period 5, when the gold extraction improved to 95.5%.

The relationship between the gold extractability and the degree of sulphur oxidation in each compartment in the integrated run is shown in Figure 3.

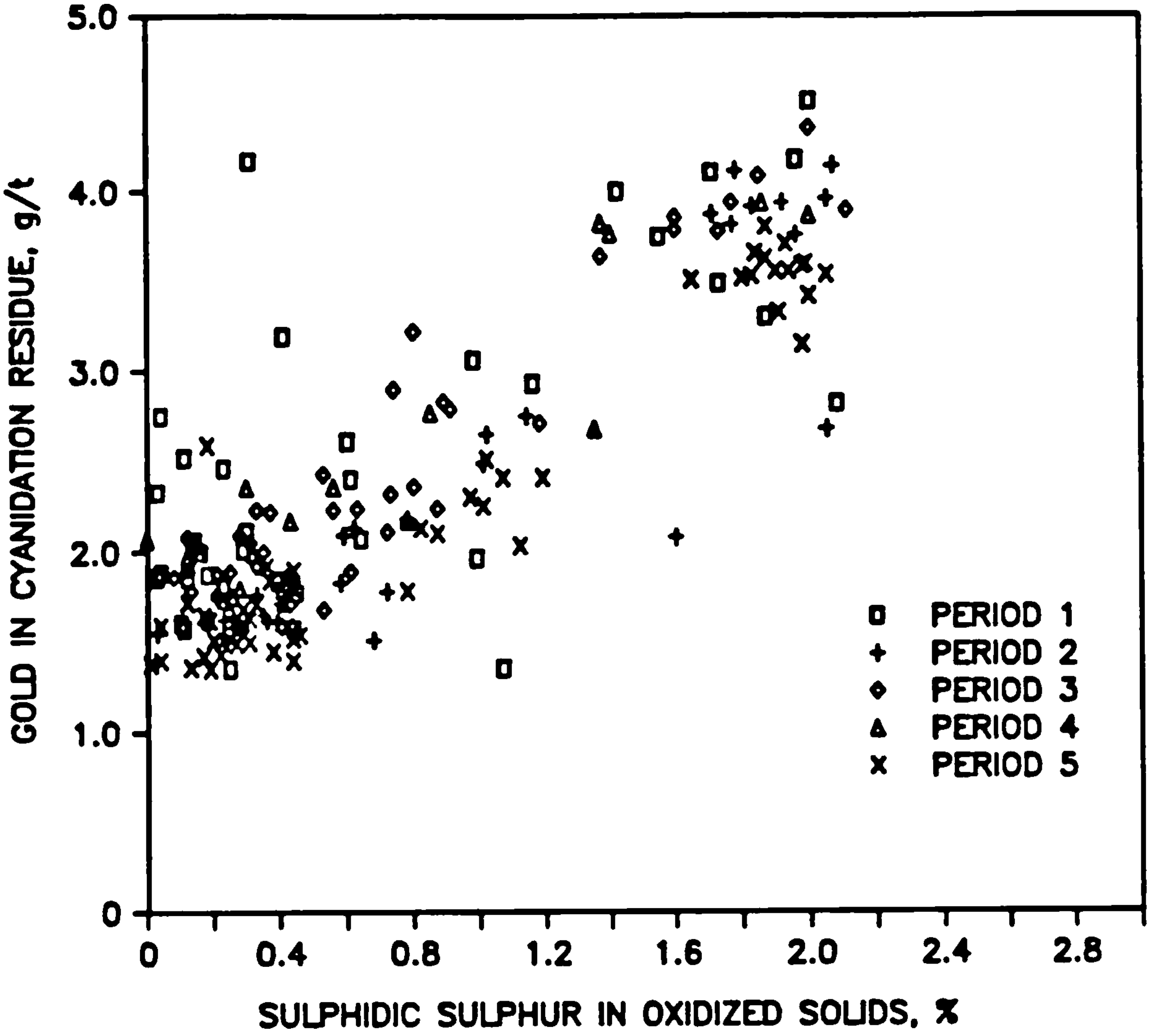

Figure 3. Gold in Amenability Residues vs Sulphidic Sulphur

The most consistent results were in Period 5, and the gold extractability was higher at a comparable degree of sulphur oxidation than during the other periods.

Oxidation Wash Circuit

The autoclave slurry was discharged into the first wash thickener feed tank, where it was diluted to 11% solids with overflow solutions from the first and second stage wash thickeners. Three quarters of the first wash thickener underflow slurry discharge was recycled to the oxidation feed slurry preparation, with the remainder proceeding to the second wash, and subsequently, to the gold recovery circuit. The first wash thickener overflow proceeded to the neutralization circuit. A summary of the wash circuit parameters and performance is given in Table VIII.

Table VIII. Countercurrent Decantation Wash Circuit Operations

Period		1	2	3	4	5
Duration, h		42	48	48	30	81
W1	Oxidation discharge, % solids	42.0	42.1	35.5	36.1	38.4
	Thickener feed, % solids	10.8	10.8	11.0	10.8	11.4
	Percol 351, g/t concentrate	92	93	94	86	93
	Underflow slurry, % solids	50.3	52.9	54.0	52.1	52.0
	Suspended solids in O/F, mg/L	133	162	167	173	164
W2	Wash ratio	15.0	14.5	14.0	12.9	13.1
	Thickener feed, % solids	6.5	6.8	7.1	7.3	7.0
	Percol 351, g/t concentrate	17	16	16	15	16
	Underflow slurry, % solids	57.2	60.3	62.5	62.6	60.2

The estimated loss of gold in the suspended solids in the first wash thickener overflow to the neutralization circuit was about 0.2% of the gold in the feed concentrate. There was no measurable segregation of gold in the first wash thickener circuit. The gold content of the second wash thickener underflow solids was variable, suggesting periods of segregation and temporary hold up of gold in the circuit.

Gold Recovery

The second wash thickener underflow slurry, containing in the range of 58 to 62% solids, was diluted to 35% solids and limed to a pH of 11.3 to 11.5 in a liming and pulp density adjustment stage, then leached in a single cyanidation stage, before proceeding to the twelve stage carbon in leach section for further gold dissolution and recovery by carbon adsorption. Full continuous gold recovery circuit operation commenced 15 h after the pressure oxidation startup, and consisted of two major periods, the first from 15 to 162 h and the second from 162 to 277 h.

The major differences between the two periods included: an increase in the solids feed rate, from 2 to 2.5 kg/h; the recycle of barren CIL solution to the pulp density adjustment stage, effectively adding a second leach stage; and an adjustment of the CIL cells to eliminate back mixing and attain effective staging. A summary of the operating parameters and conditions for the two periods is provided in Table IX.

Table IX. Continuous Gold Recovery Circuit Operations

	Run time, h	15 to 162	162 to 277
Feed	Slurry solids content, %	60.3	60.6
	Solids feed rate, kg/h	2.03	2.56
Liming and Pulp Density Adjustment	Slurry solids content, %	35.0	35.0
	Slurry temperature, °C	35	35
	Lime addition, kg CaO/t	21.6	22.3
	NaCN in recycle, kg/t	0	0.4
	Retention time, h	3.3	2.6
Leach	Fresh NaCN addition, kg/t	1.61	0.95
	Slurry pH	10.7	10.7
	Free cyanide concentration, g/L	0.80	0.63
	Retention time, h	3.3	2.6
CIL	Slurry temperature, °C	36-->28	36-->28
	Carbon advance rate, g/h	1.53	2.33
	Discharge slurry pH	10.1	10.5
	Discharge NaCN, g/L	0.55	0.50
	Retention time, h	7.35	5.63

The gold content of the circuit feed solids was initially in the 41 to 42 g/t Au range, due to residual Giant Yellowknife solids (63.5 g/t Au) in the circuits. This value decreased gradually, to the 31 to 33 g/t Au range. After about 60 h, the median gold content of the feed solids for the remainder of the run was 30.5 g/t Au.

The lime demand of the oxidized solids in the liming and pulp density adjustment stage was predominantly in the 18 to 25 kg CaO/t range, averaging 21.7 kg/t. The weight increase of the circuit feed solids during the liming reaction averaged 2.8%. The sodium cyanide addition rate was initially 1.6 kg NaCN/t oxidized solids, to maintain free cyanide levels of about 0.8 g/L and 0.6 g/L in the leach and the CIL discharge, respectively.

Gold extraction in the cyanidation stage was in the range of 77 to 92% during the first period, when there was only one stage of leaching. During the second period, the recycle of the barren CIL solution, containing 0.4 to 0.6 g/L NaCN (0.36 to 0.6 kg/t oxidized solids), resulted in the dissolution of 62 to 79% of the gold in the liming and pulp density adjustment stage. The combined addition of sodium cyanide, as fresh reagent and present in the recycle solution, was eventually reduced to 1.3 kg/t. The cyanide consumption, as determined by the free cyanide levels in the CIL discharge solutions, was in the 0.37 to 0.54 kg/t range. Gold extraction after effectively two stages of leaching was about 92.5%, ranging from 89 to 93%. The corresponding leach solids entering the CIL circuit contained 1.93 to 3.56 g/t Au. The degree of completion of gold extraction, relative to that attained by the end of the CIL circuit, was about 98.2%, ranging from 94 to 99%.

All twelve cells of the CIL circuit were operated and monitored throughout the run. The carbon concentration in each cell was 22 g/L. The target loading on the product carbon was 30 000 g/t Au. During the first period, there had been back mixing in the CIL circuit resulting in a gold loading only in the 21 000 to 24 000 g/t Au range on the lead carbon. This situation was rectified in the second period of the run, and gold loading in the 28 000 to 32 000 g/t Au range was readily achieved.

Representative gold profiles in the gold recovery circuit are provided overleaf, in Table X.

The leaching of the "extractable" gold was largely completed by about the third or fourth CIL stage. In spite of the relatively high loading of gold on the carbon, and the relatively short retention time of about 30 min/stage, the objective of <0.02 mg/L Au in solution was readily attained within nine stages.

Representative loadings of trace elements and impurities on the product carbon (from the 234 h profile) included 1210 g/t As, 7060 g/t Ca, 9 g/t Cu, 3440 g/t Fe, 108 g/t Ni and 25 g/t Zn, along with 178 g/t Ag, and 30 200 g/t Au.

Table X. Gold Recovery Circuit Gold Profiles

	Solids, g/t Au			Solutions, mg/L Au			Carbon, g/t Au		
	138 h	234 h	276 h	138 h	234 h	276 h	138 h	234 h	276 h
Feed	34.4	26.4	36.0						
Liming-PD adjustment	28.7	7.01	12.7	-	11.4	11.4			
Leach	3.74	1.93	3.55	12.7	13.8	15.8			
CIL 1	2.03	1.86	2.12	4.86	12.8	15.7	24 200	30 200	27 925
2	1.98	1.78	1.61	3.66	10.1	13.2	20 710	28 090	24 150
3	1.88	1.72	1.53	2.78	6.76	8.60	17 750	21 840	22 085
4	1.88	1.70	1.48	2.01	3.13	4.12	14 300	14 385	13 370
5	1.89	1.74	1.38	1.54	1.64	2.29	11 135	8 595	7 500
6	1.94	1.71	1.29	0.87	0.60	0.93	6 320	3 655	3 635
7	1.85	1.67	1.23	0.14	0.16	0.33	1 545	1 220	1 185
8	1.83	1.70	1.34	0.037	0.047	0.12	118	361	320
9	1.75	1.69	1.34	0.013	0.015	n.a.	37	124	n.a.
10	1.78	1.67	n.a.	0.008	0.005	n.a.	22	49	n.a.
12	1.77	1.65	n.a.	0.001	0.001	n.a.	11	13	n.a.

Loading profiles for these elements in the CIL circuit were also determined. Copper, nickel, silver, and zinc loaded to only minor or trace levels in the latter stages, and were gradually displaced by the gold. The arsenic and iron contents of the carbon were in the 1100 to 1600 g/t and 3100 to 4200 g/t range, respectively. The As:Fe weight ratio was relatively constant, at about 0.37:1.0, and was similar to that of the oxidized solids, suggesting the bulk of these species to be present as entrained leach residue. An acid wash of a sample of product carbon, containing about 28 640 g/t Au, 150 g/t Ag, 1500 g/t As, 6860 g/t Ca, 4950 g/t Fe, 113 g/t Ni and 27 g/t Zn, in dilute hydrochloric acid, removed 88% As, 94% Ca, 88% Fe, 86% Ni and 50% Zn.

A composite of loaded carbon, comprising advance carbon samples and the final carbons from cells 1, 2, and 3 was pressure stripped at 125°C. The composite carbon contained 26 120 g/t Au and 363 g/t Ag, and the carbon had not been acid washed prior to the elution. When 98.9% of the

Au and 92.8% of the Ag was stripped from the carbon, the composite strip liquor contained 1110 mg/L Au, 14.4 mg/L Ag, 47.8 mg/L As, 0.6 mg/L Cu, 71.2 mg/L Fe, 0.21 mg/L Hg, 5.4 mg/L Ni and 6.9 mg/L Zn. The mercury in the strip liquor corresponded to only about 2% of the mercury in the concentrate.

The gold accountability in the gold recovery circuit in the integrated run was 98.4%.

Neutralization Circuit

During the integrated run on the Sao Bento concentrate, the waste acidic first wash thickener overflow, in combination with Sao Bento flotation tailings, was treated sequentially with limestone, then lime. Following liquid-solid separation, the solution was recycled to the process while the solid wastes were combined with the CIL discharge slurry for deposition in a tailings structure. The chemical composition of the flotation tailings is provided in Table XI.

Table XI. Chemical Composition of Sao Bento Flotation Tailings

Sample	% Al	% As	% Ca	% CO_2	% Fe	% Mg	% SiO_2	% S
1	1.9	0.48	2.2	17.2	17.2	2.2	35.5	0.5
2	1.4	0.44	1.8	12.9	13.8	1.7	44.9	0.6

Sample 1 (75% passing 44 µm) represented hydrocyclone overflow, and was used during the first half of the run while Sample 2 (30% passing 44 µm) represented the total flotation tailings.

The neutralization circuit comprised four reaction tanks, a thickener, and a belt filter. The first wash thickener overflow solution, the flotation tailings and limestone slurry were added to tank 1, while lime slurry was added to tank 3. The pH levels in tanks 2 and 4 ranged from 4.8 to 5.8, and from 10.4 to 10.9, respectively. Addition of the simulated hydrocyclone overflow and effective flocculation and operation of the thickener yielded underflows in the range of 37 to 42% solids and enabled removal of the belt filter from the circuit. During the second half of the run, total flotation tailings were added and the thickener underflow solids content increased to between 51 and 61%. Limestone and lime consumptions during the latter part of the run averaged 1.08 and 0.19 kg/h, respectively.

The possibility of reducing the limestone requirements, by utilizing more of the neutralizing capacity of the flotation tailings was evaluated in a subsequent integrated run, utilizing arsenopyrite concentrate from Giant Yellowknife Mines in the oxidation circuit, and Sao Bento flotation tailings in the neutralization circuit. An additional reaction tank was added to the neutralization circuit, where flotation tailings were reacted with the acidic wash thickener overflow solution, to utilize a portion of the acid consuming content of the tailings prior to addition of the limestone. Although this practice did in fact reduce the limestone requirements, the reaction of a portion of the acid with the siderite in the tailings generated ferrous sulphate, which in turn required further oxidation and increased the lime addition requirements.

The median compositions of the acid waste solutions and of the neutralization thickener overflow produced during the two runs are provided in Table XII.

Table XII. Neutralization Circuit Feed and Product Liquors

Element	Feed Assays, mg/L		Product Liquor, mg/L	
	Run 1	Run 2	Run 1	Run 2
Al	356	1543	n.a.	n.a.
As	398	442	0.10	0.09
Cd	5.1	8.4	<0.005	<0.005
Cr	n.a.	3.0	0.008	0.007
Co	2.0	30	<0.01	0.01
Cu	16.7	113	0.02	0.03
Fe	4000	4610	0.08	<0.01
Pb	n.a.	0.55	<0.05	<0.05
Hg	n.a.		<0.01	<0.01
Mg	687	1360	0.9	<0.1
Mn	50	58	0.02	<0.01
Ni	12.6	52	0.02	0.03
Se	n.a.	n.a.	<0.01	n.a.
Ag	n.a.	n.a.	0.02	n.a.
Zn	26.6	328	0.01	0.01

* n.a. - not analyzed

The impurities of environmental concern were readily removed to trace levels.

Tailings Disposal

The subaerial tailings deposition technique was tested for the disposal of the neutralization circuit waste solids and the CIL discharge solids. Neutralization thickener underflow slurry, and settled CIL discharge slurry were combined in a weight ratio of 4:1. After adjustment of the slurry blend to about 50% solids, successive layers of the blend tailings were laid down in either of two specially designed tailings boxes, one containing a base layer of compacted, graded sand, and the other the coarse (plus 88 μm) fraction of Sao Bento flotation tailings, simulating hydrocyclone underflow solids.

The addition of layers to the tailings boxes was conducted over a nine week period. After deposition, each ~10 cm layer of slurry was allowed to drain and partially dry by natural evaporation, before the next layer was put down. After 11 to 21 days, the surface layer cracked and formed a solid deposit containing 25 to 35% moisture. Only about 5% of the input water was recovered as runoff or seepage water. The remainder was lost by evaporation. The study confirmed that the barren solids are well suited to treatment by the subaerial deposition technique.

COMMERCIAL PLANT FLOWSHEET

The results of this very extensive experimental programme provided the basis for the design of the commercial plant of Sao Bento Mineracao. Sherritt undertook the process design of the pressure oxidation, counter-current decantation washing and neutralization sections as well as the detailed design of the autoclaves whilst General Mining Union Corporation provided the process design of the milling and gravity, flotation, carbon in leach and electrowinning and smelthouse sections.

The plant is presented schematically in Figure 4 and is described below.

Ore is delivered to a run of mine grinding circuit comprising a fully autogenous open circuit primary mill followed by a secondary ball mill operating in closed circuit with a gravity concentration plant to remove free gold and a cyclone classifier.

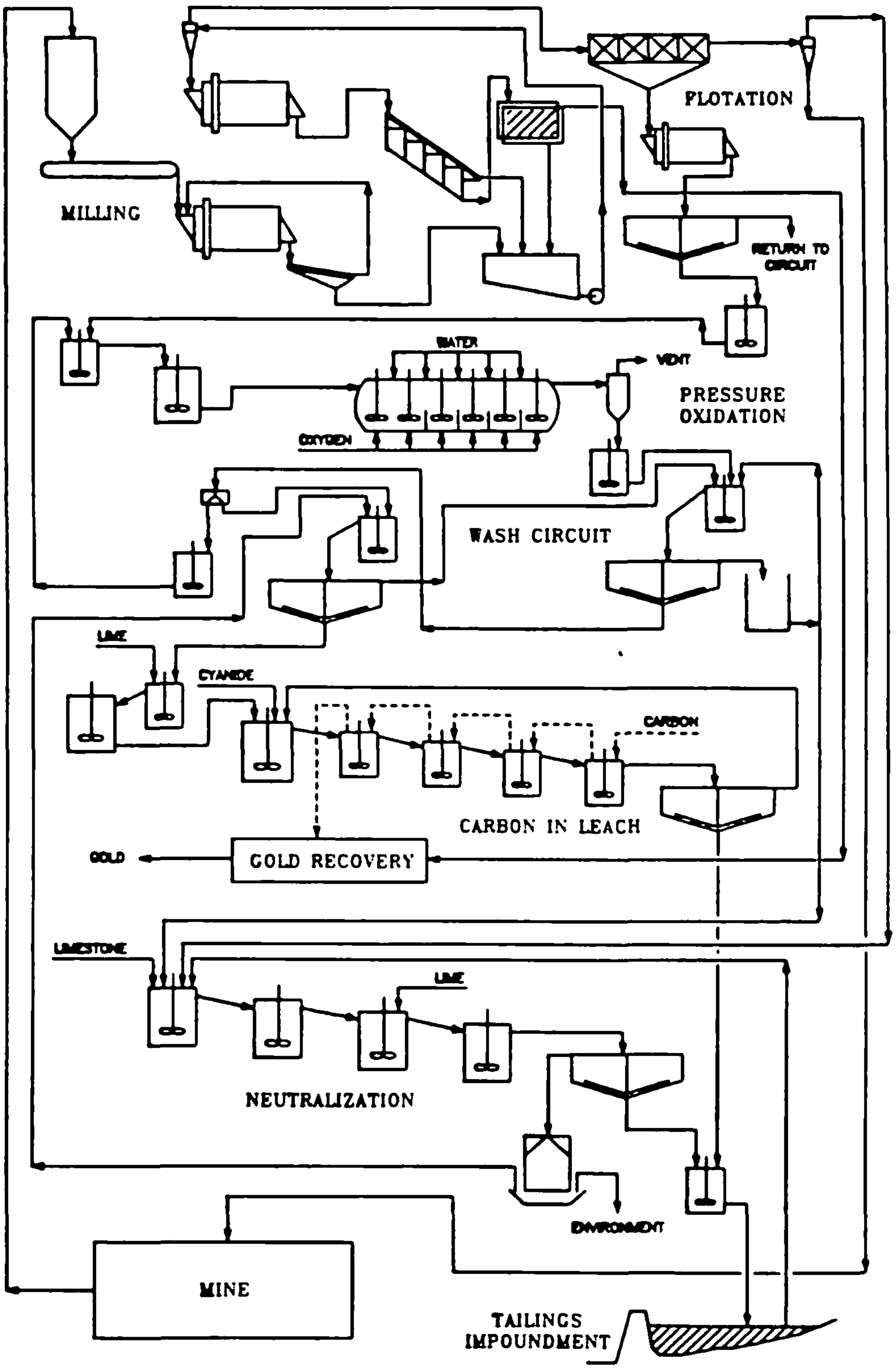
FLOTATION
MILLING
RETURN TO CIRCUIT
WATER
VENT
PRESSURE OXIDATION
OXYGEN
WASH CIRCUIT
LIME
CYANIDE
CARBON
CARBON IN LEACH
GOLD
GOLD RECOVERY
LIMESTONE
LIME
NEUTRALIZATION
ENVIRONMENT
MINE
TAILINGS IMPOUNDMENT

Figure 4. Sao Bento Refractory Gold Process

The cyclone overflow from the milling circuit forms the feed to a conventional sulphide flotation circuit. The flotation tailings are hydrocycloned, with the hydrocyclone underflow being recycled to the underground mine to be used as backfill. The overflow is delivered to the waste control section. The flotation concentrate is reground and thickened. The thickened pulp is stored and blended.

The concentrate slurry is contacted with acidic slurry derived from a countercurrent decantation washing circuit. The concentrate contains carbonate which reacts in this acidic slurry evolving carbon dioxide. The slurry is then pumped into one of two 5 compartment 6 agitator autoclaves. Oxidation is conducted at elevated temperature and elevated pressure under oxygen. The autoclave discharge slurry is flashed to release excess heat. The flashed slurry is diluted with recycled thickener overflow solution as well as wash solution and then flocculated and thickened. The thickener underflow is stored and distributed to the second wash thickener or recycled to the oxidation circuit. Makeup water for the wash circuit is recovered from a cooling pond.

The washed solids are neutralized to pH >10.5 with lime and then cyanide leached and subjected to carbon in leach treatment in a multistage circuit. The carbon, containing the gold adsorbed in the carbon in leach circuit, is delivered to a gold elution section. The gold is eluted from the carbon using the Zadra system and is recovered by electrowinning. The wire wool cathodes are then chemically refined to produce a final product of 99.9 fineness. The barren slurry discharged from the carbon in leach circuit is thickened and the thickener overflow is recycled to the circuit.

The overflow solution from the first wash thickener is treated sequentially with flotation tailings and limestone then lime to precipitate the metals and the sulphate that were derived from the concentrate. This slurry is thickened and the metal free water is cooled in a cooling pond for recycle to the washing circuit. Excess water is discharged to a monitoring pond and after certification is discharged to the environment. The thickener underflow is combined with the thickened carbon in leach discharge residue and pumped to a tailings dam which uses a subaerial deposition technique. Solution bleeding from the tailings deposit is recycled to the waste control circuit.

The plant commenced operations in November 1986 and reached its design capacity of 20 000 tons per month of ore by April 1987. Since then, it has been operating consistently in excess of 25 000 tons per month. A description of the plant and its operation will be the subject of a future paper.

Acknowledgements

The authors wish to thank the managements of Sao Bento Mineracao, General Mining Union Corporation Limited, and Sherritt Gordon Mines Limited for permission to present this paper and for their enthusiasm and support for this approach to process development. The significant contributions of the professional and technical staff of Sherritt's Process Research Group are also duly acknowledged.

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STARTUP OF THE SHERRITT PRESSURE OXIDATION PROCESS AT SAO BENTO

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1988-256

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ABSTRACT

The Sao Bento gold plant, located in the Minas Gerais province of Brazil was started up in late 1986. This plant uses the Sherritt Gordon Pressure Oxidation Process to treat refractory concentrates containing significant quantities of arsenopyrite. This paper briefly describes the process, then highlights startup problems and solutions, followed by a summary of current operations.

INTRODUCTION

The Sao Bento gold deposit is located in the Minas Gerais province of Brazil approximately 500 km north of Rio de Janeiro. The major sulphidic minerals in the ore are arsenopyrite, pyrite and pyrrhotite, and a significant portion of the gold content is refractory to recovery by conventional treatment. Following evaluation of several alternatives for processing of a flotation concentrate, the Sherritt process utilizing pressure oxidation was chosen for the commercial plant.

An oxidation testwork contract was placed with the Sherritt Research Centre in May 1983 (1). This consisted of batch tests, followed by testing of the various unit operations and subsequently a continuous fully integrated run in a miniplant. This work was conducted between November 1983 and mid February 1984.

A process design package, and detailed specifications for specialized equipment, were then developed from the miniplant testwork data. This enabled an engineering contractor to complete the design for a grass roots plant to treat nominally 240 t/d of concentrate. The process design package was started in October, 1984 and gold was produced in January, 1987.

This paper focuses on the pressure oxidation and utilities sections of the project, describing the process, startup problems and solutions, followed by a brief summary of current operations. A more detailed appraisal of the whole project will be the subject of a future paper.

PROJECT OVERVIEW

The Sao Bento complex is a fully integrated operation consisting of an underground mine, mill and concentrator, pressure oxidation, residue washing, waste treatment and gold recovery including production of fine gold. The facility also includes all necessary infrastructures including utilities such as water supply and treatment, steam plant and air systems and a modern analytical laboratory. Oxygen is purchased across the fence from a dedicated tonnage oxygen plant. Figure 1 shows the schematic of the process and highlights those processing sections which are discussed in more detail in the following sections.

PROCESS DESCRIPTION

FEED PREPARATION

Reground concentrate (see Table 1) is delivered from the flotation circuit as a slurry containing 65 wt% solids. It is metered into the feed preparation reaction tank where it is mixed with slurry recycled from the first wash thickener underflow tank. The flow and density of the reground concentrate slurry are measured and integrated to enable mass flow control. The recycled slurry contains sulphuric acid which reacts with some of the magnesium and calcium carbonates present in the concentrate, liberating carbon dioxide. After addition of lignosol solution, slurry is transferred from the reaction tank to the oxidation feed storage tank which provides surge volume for the oxidation circuit and additional residence time for the reaction of sulphuric acid and carbonates.

TABLE 1. CHEMICAL COMPOSITION OF DESIGN CONCENTRATE

Component	Al	As	Ca	CO_2	Fe	Mg	S
Analysis (%)	1.6	9.9	1.0	6.0	34.0	1.1	18.7

PRESSURE OXIDATION

Slurry is pumped from the feed storage tank to each of two parallel autoclave trains. Each train consists of an autoclave feed pump, autoclave and flash tank. Pretreated slurry is pumped at a set rate, into the oxidation autoclaves using positive displacement diaphragm pumps. Oxygen is fed into the autoclaves to oxidize the sulphides contained in the concentrate. Clarified river water is also pumped into the autoclaves to maintain the heat balance. Pressure oxidation is carried out for a nominal 120 minutes in two 5 compartment horizontal vessels 3.5 m diameter and 19.0 m long each equipped with six agitators. The lead and brick lined autoclaves operate at 190°C and a pressure of 1 560 kPa (ga). Slurry from the last compartment of each autoclave is discharged through a choke to separate lead and brick lined flash tanks operating at atmospheric pressure. Steam from the flash tanks, after passing through a demister, is vented to the atmosphere and the discharge slurry drains to a common seal tank located beneath each flash tank. Autoclave vent gases, after passing through a knock-out pot, are vented to the flash tanks where any gaseous sulphur present in the gas is condensed.

RESIDUE WASHING

The flash tank discharge slurry is diluted by a flow of second thickener overflow solution to the flash tanks and seal tank. The slurry is then pumped from the seal tank to the first wash repulp tank, where it is further diluted with recycled first wash thickener overflow solution. The repulped slurry then overflows to the 25 m diameter first wash thickener after the addition of diluted flocculant solution.

The first wash thickener underflow is pumped to the underflow splitter box. The majority (80%) of the underflow slurry is directed to the first wash thickener underflow tank. Slurry from the underflow tank is recycled to the feed preparation reaction tank at a controlled rate. The remainder of the first wash thickener underflow is directed from the splitter box to the second wash repulp tank, where it is diluted with cooled recycle water from the cooling pond. This slurry overflows to the 25 m diameter second wash thickener after diluted flocculant solution is added to it. A portion of overflow solution from the first wash thickener is transferred to the neutralization circuit and a portion is recycled to the first wash repulp tank. The second wash thickener underflow is pumped to the gold recovery circuit. The second wash thickener overflow is recycled to the flash tanks and the seal tank.

NEUTRALIZATION

The first wash thickener overflow solution metered to this section of the plant is partially neutralized in agitated cascading tanks in series. Flotation tailings with neutralizing potential from the milling circuit can be added to the first limestone reaction tank. Limestone slurry is added from a splitter box to the first tank raising the pH of the slurry to about 4.5. The slurry overflows to a second tank and third tank and then to a smaller, well agitated, lime addition tank where lime slurry is added from a splitter box. The slurry then overflows to two agitated lime reactions tanks, which ensure adequate retention time for the precipitation of the remaining metals in solution. The neutralized slurry, at a pH of 11, is advanced from the second lime reaction tank to the sludge thickener. The pH is monitored after the first limestone reaction tank and after the lime addition tank.

The sludge thickener underflow is pumped to the sludge thickener underflow tank, where it is combined with the tailings from gold recovery and advanced to the tailings dam. The neutralized sludge thickener overflow solution is pumped to a cooling tower for cooling and recycle to the process.

LIGNOSOL SOLUTION MAKE-UP

The daily quantity of 25% sodium lignosulphonate solution is made up each day with potable water. The powder is transferred manually to the solution mixer where circulating potable water (or solution) wets the powder and transfers it to an agitated make-up tank. The solution is transferred daily to the feed tank where it is metered to the feed preparation reaction tank.

FLOCCULANT ADDITION

The flocculant solutions for both the wash circuit thickeners and the neutralization sludge thickener are made up in separate, but similarly designed, circuits. In each case, the daily requirement of flocculant is mixed with process water in an agitated mix tank. The solution is transferred by gravity once a day to a feed tank, where it is metered to the respective areas.

SLAKED LIME MAKE-UP AND LIMESTONE MAKE-UP

The pebble lime and limestone, in separate slurry makeup circuits, are conveyed from storage bins to the ball mills, at set rates on weigh belts. Cooled recycle water from the cooling tower is metered to each mill to obtain the desired pulp density. Each ball mill discharge slurry is pumped to a storage tank from which the respective lime and limestone slurries are pumped to the required process area. The lime slurry is pumped to a splitter box in neutralization and also batch transferred 4 hours per day to the lime slurry storage tank in gold recovery. Excess lime slurry from the splitter box is returned to the storage tank. The limestone slurry is pumped to a splitter box in neutralization with excess slurry returned to the storage tank.

UTILITIES

The main utility systems include water supply and treatment, steam generation, plant air systems and electrical supply.

WATER SUPPLY AND TREATMENT

Water is obtained from the Conceicao River which runs in the valley 700 feet below the plant site. Facilities at the river include river intake, presettler, gravity sand filters and a pumping station. The filtered water is pumped to a water reservoir near the plant site.

A portion of the water is treated in water softeners for production of boiler feed water and autoclave agitator seal water.

STEAM GENERATION

Steam at 2 200 kPa is required to heat the autoclave contents during startup. Small quantities are required during operations to assist in autoclave sampling and for heat balance purposes. The Gold Recovery area has a small requirement for low pressure steam.

Due to the large turndown requirement and also for reliability, two electric steam boilers were provided.

PLANT AIR SYSTEMS

Compressed air for the plant is supplied by two screw-type compressors which have a capacity of 27 m^3/minute at a pressure of 700 kPa.

A portion of the air is dried in a refrigeration type drier for instrumentation purposes.

ELECTRICAL SUPPLY

Power is obtained from the State Electricity Authority (CEMIG) 230 kV network via a switch yard situated 12.5 km from the mine.

Standard three phase supply at the mine is 440 V. Emergency power is automatically available for select equipment via a diesel generator in the event of a power failure.

In addition, battery power for the instrumentation control system is available from a UPS (uninterrupted power supply) system.

START-UP PROBLEMS AND SOLUTIONS

The pressure oxidation section of the plant first received concentrate at the end of November 1986. The first gold was produced in January 1987 and full production was reached in April 1987. This was a remarkably rapid commissioning period for what is a sophisticated and complex plant. This can be attributed to timely recruitment and extensive training of the senior Brazilian operations staff and a carefully planned startup.

The training of the operational staff consisted of five elements:

1. Three months of training at Gencor operations in South Africa down to the level of section head (foreman).
2. Flow sheet and operating manual training down to the level of operator.
3. Lectures and instruction by Sherritt.
4. Computer simulation training down to shift supervisor level.
5. On the job training at all levels.

This training commenced some six months before the startup of the plant.

The startup team was divided into four basic groups:

- a technical group comprising mainly expatriates;
- the plant operations group under the metallurgical manager who was seconded from Gencor;
- the maintenance group under the plant engineer; and
- the assay laboratory.

These groups were co-ordinated by means of a daily meeting chaired by a Gencor consulting metallurgist.

The expatriates of the technical group comprised 12 engineers and operational specialists from both Gencor and Sherritt who were on site in varying numbers from two months before startup to four months afterwards.

The startup philosophy used was that actual starting and running of the plant equipment was the responsibility of the operations group.

The technical group provided shift technical supervision and control of the plant. Changes to operating procedures were, however, done through the operations group. This philosophy was used to ensure that the operations group was placed on the learning curve from day one and not just handed over a fully operating plant.

A major negative aspect of this approach was, of course, the language barrier, but steps were taken to alleviate this problem.

A vital aspect of the startup was the early commissioning of the analytical laboratory. This commenced functioning in July 1986 so that by the time the plant started, the laboratory was fully operation and vital analytical results were available within hours of submission of samples when required.

The startup was not without its problems and many mechanical breakdowns occurred in the early months.

Mechanical problems, associated with a sophisticated technology, new to the country and in a remote location inevitably were magnified by shortages of materials and spare parts. Laws imposed by the Brazilian government restrict the importation of materials such as alloy piping, pipe fittings and valves. For example, replacement valves which would have taken days to procure in North America, could require months, if allowed at all, to be imported into Brazil. If a Brazilian company stated that it can manufacture the item, then importation was not permitted. Often these items were then manufactured by fabricators with little previous experience. This could result in design faults and subsequent failure.

During startup, the operations personnel quickly devised means to operate without certain control valves and the maintenance staff became very adept at salvaging material and fabricating it to satisfy a need. It is a tribute to the Sao Bento staff and startup supervisors that the plant reached and sustained design rates within four months of feed introduction.

Some specific problems experienced during startup and their solutions are described below.

UTILITIES

As in many startups, the utilities and auxilliary systems provided a disproportionate number of problems. In addition to the usual problems of leaking fittings and flanges, many spurious electrical trips and computer outages were encountered.

The most serious utility problem concerned the water supply to the plant. The silt load in the river increased dramatically over design. This was due to a dredging operation which started upsteam of the water intake, about the same time that the commissioning started. Commissioning was also coincident with naturally high silt loads in the river due to the runoff during the rainy season. The heavy silt load problem was compounded due to the failure of the lining in the water storage reservoir, which provides for some solids settlement. The heavy silt load limited the quantity of suitable water available for commissioning purposes. During this period, the water softeners were used to clarify the river water in order to progress with plant commissioning. The lack of the desired quantity of clean water extended the commissioning and startup period by several days. This problem was alleviated by repair of the liner in the short term, and the addition of a presettler in the long term.

FEED PREPARATION

Agitators proved to be a major mechanical problem throughout all areas of the plant. Two mechanical design faults became evident immediately. The gearbox design was faulty, causing loss of lubricating oil which entered the process tankage. The first fill of concentrate slurry in the feed storage tank had, in fact, to be dumped to the tailings dam due to oil contamination. The second fault concerned the attachment of rubber coated impeller blades to rubber coated shafts. The blades were bolted to the shaft stubs. The rubber to rubber contact resulted in movement and the bolts worked loose or sheared, dropping the blades. The gearboxes on an individual basis, had to be returned to the manufacturer for correction and each blade to stub contact area was reworked to eliminate the rubber to rubber contact.

Air sparging was used for agitation while the gearbox was being repaired, allowing operations to continue.

Pitting of the internal surface of the carbon steel concentrate slurry storage tanks occurred soon after commissioning. These tanks have since been epoxy coated.

AUTOCLAVE CIRCUIT

An error in the level control setting on the autoclave vent knockout pots resulted in slurry carry over into the gas vent lines from the autoclave. This caused early failure of the pressure control valves. Replacement valves have proven to be difficult to obtain in Brazil for reasons stated earlier. The plant has operated with restriction orifices in the vent line since failure of the valves.

The PLC based computer control system experienced many failures during the early commissioning and startup periods. These failures were associated with programming errors, electrical faults and a warm control room. Air conditioning for the control room proved to be imperative. The original plan called for the vendor representative to restart the computer system. This was found to be impractical, so the control operators and supervisory staff were trained to reset the PLC and computers. This was very successful and certainly reduced downtime due to this problem.

Several autoclave agitator seal failures were experienced. This problem was corrected after it was discovered that the seals were incorrectly assembled.

The Geho charge pumps ball check valves failed prematurely due to a fabrication problem with the protective cover.

Some corrosion and failure of internal metal components within the autoclave occurred during the first months of operation. These problems were caused by several factors. In some cases, the corrosion occurred on materials which were downgraded due to shortages during construction. Oxygen sparger supports and cell end baffles proved to be under designed, and corrosion has been experienced at the gas liquid interface on the dip pipes, particularly the steam sparger. Changes have been incorporated and better than design on stream time is now achieved.

RESIDUE WASH CIRCUIT

No unusual problems were encountered in the residue wash circuit. Once experience was gained by the operators, this circuit performed as designed.

WASTE TREATMENT

The waste treatment circuit initially experienced difficulties due to the agitator problems mentioned earlier. By careful control of pH, soluble arsenic levels are consistently reduced to below the detection limit of 0.05 mg/L.

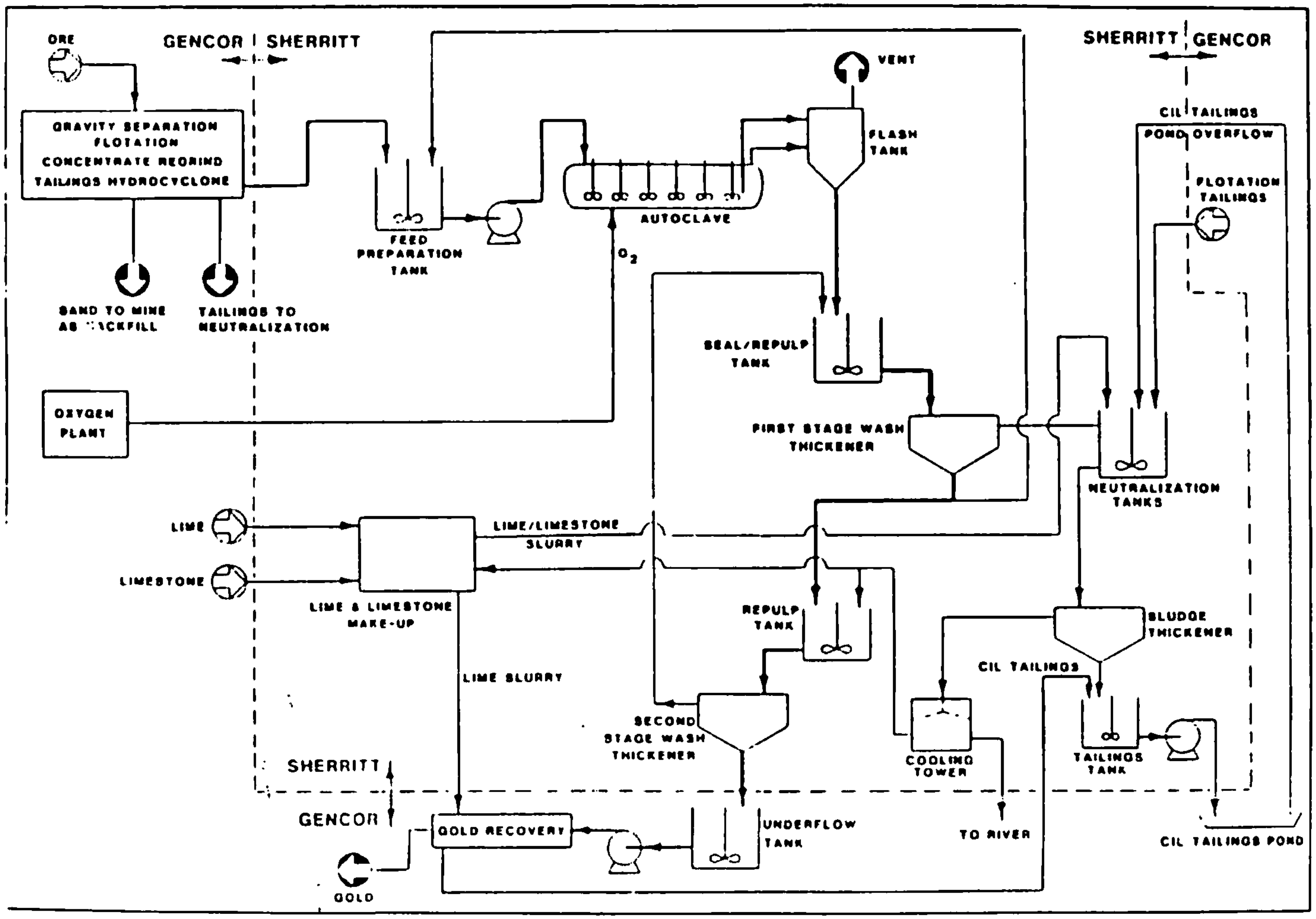
CURRENT OPERATIONS AND CONCLUSIONS

The process has responded as predicted from the laboratory and pilot plant tests, and process guarantees provided by Sherritt have been met. Since reaching full production in April 1987, the plant has consistently performed at its design rate.

Oxygen utilization efficiency has been slightly below design. This is mainly due to a higher than projected carbonate content of the concentrate resulting from a "harder pull" in the flotation circuit. This mode of operation leads to improved gold recovery in the flotation circuit but at the expense of autoclave efficiency. Planned changes in the agitation can be expected to improve oxygen transfer efficiency. Gold Recoveries should also improve when this change is made since higher sulphide oxidation should also be achieved.

Scaling has occurred in the neutralization tanks, the second wash thickener and in certain pipelines. This is, however, completely manageable. The descaling has only had to be done once in the 20 months of operation. No loss of production has occurred.

FIGURE 1. SHERRITT PRESSURE OXIDATION PROCESS


ORE
GENCOR | SHERRITT
GRAVITY SEPARATION
FLOTATION
CONCENTRATE REGRIND
TAILINGS HYDROCYCLONE
SAND TO MINE AS BACKFILL
TAILINGS TO NEUTRALIZATION
OXYGEN PLANT
FEED PREPARATION TANK
AUTOCLAVE
O_2
VENT
FLASH TANK
SEAL/REPULP TANK
FIRST STAGE WASH THICKENER
SHERRITT | GENCOR
CIL TAILINGS
POND OVERFLOW
FLOTATION TAILINGS
NEUTRALIZATION TANKS
LIME
LIMESTONE
LIME/LIMESTONE SLURRY
LIME & LIMESTONE MAKE-UP
LIME SLURRY
REPULP TANK
SLUDGE THICKENER
CIL TAILINGS
SECOND STAGE WASH THICKENER
COOLING TOWER
TAILINGS TANK
SHERRITT
GENCOR
GOLD RECOVERY
UNDERFLOW TANK
TO RIVER
CIL TAILINGS POND
GOLD

ACKNOWLEDGEMENTS

We thank Sao Bento Mineracao S.A., General Mining Union Corporation Limited and Sherritt Gordon Limited for permission to publish this paper.

The dedication and efforts of the operating and maintenance personnel at Sao Bento and those involved in providing around the clock startup assistance from Gencor and Sherritt are also gratefully acknowledged.

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Presented at Projects '88, 18th Hydrometallurgical Meeting held in Edmonton, Alberta, Canada May 9 to 11, 1988.

REFRACTORY GOLD: THE ROLE OF PRESSURE OXIDATION

by

D. Robert Weir
and
Roman M.G.S. Berezowsky

Presented at Gold 100, the International Conference on Gold held in Johannesburg September 15-19, 1986. Sponsored by the South African Institute of Mining and Metallurgy. Published in GOLD 100. Proceedings of the International Conference on Gold. Volume 2: Extractive Metallurgy of Gold.

1986-236

REFRACTORY GOLD :
THE ROLE OF PRESSURE OXIDATION

by

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SYNOPSIS

Many procedures, such as intense grinding, preaeration, chlorine oxidation, flotation followed by roasting, as well as others, have been commercialized for the recovery of gold from refractory feedstocks. Some sulphidic ores and concentrates do not respond economically (cost and recovery) to these procedures. Pressure oxidation can virtually always render gold bearing materials amenable to cyanidation and is one of the more recently developed techniques for treatment of refractory gold materials. A variety of ores and concentrates have been subjected to integrated miniplant testing and process demonstration by Sherritt. Some factors affecting the selection of pressure oxidation in gold recovery processing are discussed.

INTRODUCTION

Gold ores are classified as refractory when a significant portion of the gold cannot be extracted efficiently by conventional means, even after fine grinding. A common cause of refractoriness is dissemination of fine grained or submicroscopic gold within sulphide minerals, usually arsenopyrite or pyrite. Such ores require a pretreatment to alter or destroy the sulphide matrix to render the gold amenable to recovery.

The traditional method for the treatment of complex sulphide ores involves an oxidative roast of a flotation concentrate. Not all ores respond favorably to flotation, and the attainment of the sulphur grade required for autogenous roasting is often at the expense of gold recovery. Roaster operation and gold recovery from the calcine are sensitive to the antimony and lead contents in the concentrate. Furthermore, the generation of sulphur, arsenic, antimony, and mercury containing gases is most undesirable from an environmental standpoint and, unless there is an attractive market for sulphuric acid, control is extremely expensive.

The significant increase in exploration activity during the past decade in response to the dramatic rise in the price of gold has resulted in not only the development of many new conventional orebodies, but also in the delineation and development of more refractory orebodies. This, together with the disadvantages of roasting, has spurred the development of alternative and more efficient processes.

Most of this activity has been on hydrometallurgical oxidation processes, and includes atmospheric oxidation in alkaline medium with chlorine, atmospheric biological oxidation in acid medium, and pressure oxidation, in both alkaline and acid systems(1). The alkaline chlorination process, as practised at Carlin and Jerritt Canyon, is effective where the refractoriness is due to the presence of carbonaceous matter with "preg robbing" characteristics and where gold is disseminated in readily oxidizable sulphides. Most sulphidic ores, however, require extensive oxidation and the cost of chlorine becomes prohibitive.

The application of biologically assisted oxidation for the pretreatment of refractory sulphidic gold ores and concentrates is a relatively new concept which is presently being actively pursued at laboratory, pilot and semicommercial scales(2,3,4). Many of the earlier concerns, such as long retention times and sensitivity of the microorganisms to various influences are apparently being overcome. Biological oxidation, by virtue of its mild nature, relative to roasting or pressure oxidation, offers a certain degree of selectivity in the oxidation of arsenopyrite relative to pyrite. This may offer the potential of acceptable gold liberation from arsenopyrite-pyrite ores with only a partial oxidation of the sulphides if the refractory gold is largely associated with the arsenopyrite. Many sulphidic feedstocks, however, require extensive sulphur oxidation for effective gold liberation. Long residence times, high power costs for agitation, and the need for effective heat removal are some of the constraints which may limit the applicability of the biological oxidation approach.

Pressure oxidation is conducted at relatively high temperatures, 180 to 210°C, achieving essentially complete oxidation of the sulphides within 1 to 3 h. Virtually always the gold is rendered amenable to high recovery by cyanidation. The responses of over 50 ores and concentrates are depicted in Figure 1.

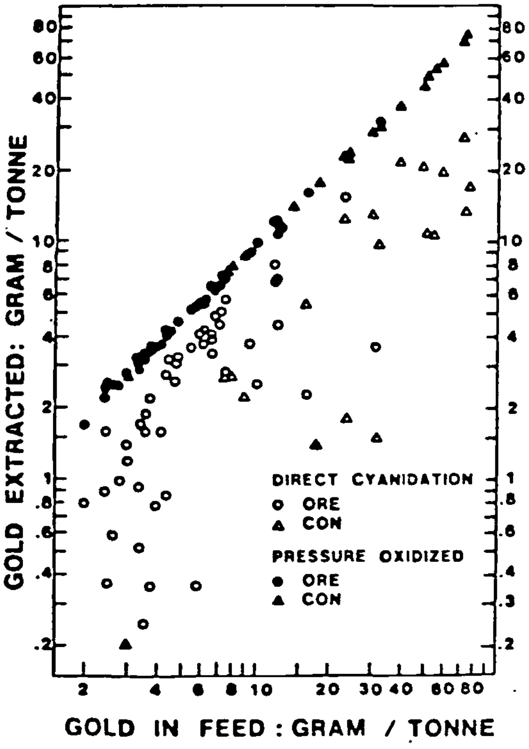

Figure 1 Effect of Pressure Oxidation on Gold Extraction

The feed stocks varied in their degree of refractoriness. After pressure oxidation, consistent recovery rates, affected modestly by the head grade, were achieved(5,6). Although the concept of pressure oxidation pretreatment is not new, dating back to the mid fifties, the renewed interest due to high gold prices has gained considerable impetus with Homestake's decision to employ pressure oxidation at their McLaughlin property(7,8). Several reviews and publications have become available on pressure oxidation(9,10,11), and a recent study of various process options concluded that pressure oxidation is the prime choice for the treatment of refractory gold ores and concentrates(12).

During the past five years, Sherritt has evaluated the applicability of pressure oxidation on a variety of refractory gold ores and concentrates. The studies have ranged from batch tests to completely integrated miniplant testing consisting of pressure oxidation, precious metals recovery, and waste control. Data derived from these studies have been used in the design and construction of Homestake's McLaughlin gold plant in California, which has recently come on stream. The testwork also led to the design of the process flowsheets for Gencor's Sao Bento Gold Project in Brazil, which is scheduled to come on stream in late 1986(13), and for the Porgera Gold Project in Papua New Guinea(14,15).

Process Chemistry

An account of the pressure oxidation chemistry has been given in previous publications(1,5,10). The major reactions in the elevated temperature oxidation of sulphides in an acid medium involve total oxidation of the sulphidic sulphur to the sulphate form.

$$2FeS_2 + 7\ O_2 + 2H_2O \dashrightarrow 2FeSO_4 + 2H_2SO_4$$

$$4FeAsS + 11\ O_2 + 2H_2O \dashrightarrow 4HAsO_2 + 4FeSO_4$$

$$MS + 2\ O_2 \dashrightarrow MSO_4$$

where M = Pb, Zn, Cu, Ni, Co, etc.

The divalent iron and trivalent arsenic are further oxidized, to their respective ferric and arsenate states. Consumption of some of the acid by gangue components, and the elevated temperature operation promote appreciable hydrolysis and precipitation of the arsenate and ferric iron, as ferric arsenate, hematite, basic ferric sulphates, and jarosites. The major prerequisites for effective oxidation of the sulphides include sufficient acidity to maintain adequate iron levels in solution to promote the oxidation, and operation under conditions which minimize the incidence of elemental sulphur formation.

Flowsheet Description

The choice of processing ores or concentrates is governed by a number of factors. For illustrative purposes, a conceptual flowsheet is given in Figure 2. Since many feedstocks may contain appreciable levels of carbonates, the feedstock is preconditioned with an acidic recycle stream to evolve as much carbon dioxide as possible prior to autoclaving. This serves to improve subsequent oxygen utilization by virtue of minimizing losses of oxygen in carbon dioxide bleed gases from the autoclave. The pretreated material then is oxidized in the autoclave with oxygen, using the heat generated by the sulphide oxidation to sustain the reaction. Water may be added, as required, for temperature control. The autoclave discharge slurry is cooled by flashing, and is subjected to a

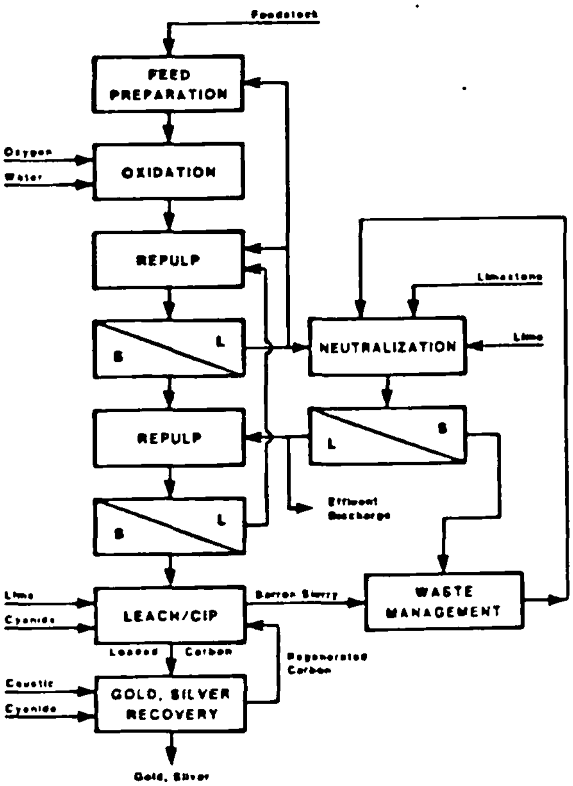
Feedstock
FEED PREPARATION
Oxygen
Water
OXIDATION
REPULP
S
L
NEUTRALIZATION
Limestone
Lime
REPULP
L
S
Effluent Discharge
S
L
Lime
Cyanide
LEACH/CIP
Barren Slurry
WASTE MANAGEMENT
Loaded Carbon
Regenerated Carbon
Caustic
Cyanide
GOLD, SILVER RECOVERY
Gold, Silver

Figure 2 Conceptual Flowsheet for Refractory Feedstocks

liquid-solid separation to remove a portion of the acid, dissolved arsenic, and metal sulphates. Washing of the oxidized pulp prior to cyanidation is generally advisable. This serves to remove not only cyanicides, such as copper and zinc, released during the pressure oxidation, but also aluminum, iron and magnesium which would otherwise precipitate as slimy hydroxides, increasing the slurry viscosity, the possibility of gold losses through adsorption by the slimes during cyanidation, and fouling of the activated carbon.

A portion of the acidic liquor is recycled to the pretreatment stage while the remainder is treated in a neutralization circuit to neutralize the acid and precipitate the arsenic, metals and associated sulphate. Water is reclaimed for recycle to the wash circuit. Since flotation tailings generally contain appreciable carbonate levels, they are often advantageously added to the neutralization circuit to reduce limestone and lime requirements.

The washed, oxidized solids proceed to a gold recovery operation, for adjustment of the pH with lime and recovery of the gold, generally by cyanidation and precipitation, cyanidation and carbon in pulp (CIP) processing, or carbon in leach (CIL) processing. The sludge from the neutralization circuit, preferably dewatered by thickening, and the barren slurry from the gold recovery operation are combined for deposition in a tailings impoundment area. Runoff water from the tailings area normally is recycled.

Applicability of Pressure Oxidation

A favorable response of a refractory gold ore to pressure oxidation may not necessarily justify the pretreatment. The applicability of pressure oxidation depends on numerous factors including ore reserves, sulphur grade, quantity of refractory gold relative to the sulphur level, nature of the gold association with the sulphides, power costs, lime costs, etc. The possibility of effecting a high degree of gold liberation with only partial oxidation of the total sulphides may favour consideration of biological oxidation. Variability of the ore grade and of the degree of refractoriness are also important considerations. It is interesting that in recent evaluations of pressure oxidation by two companies(7,16), Homestake selected pressure oxidation while Getty did not, in spite of similar ore gold grades and responses to cyanidation without pretreatment. In the latter case, the refractory ore constitutes only a portion of the total orebody and the caustic oxidation approach evaluated by Getty may have liberated some environmentally objectionable species. In the case of Homestake, acid pressure oxidation, which was effective for the wide range of ores tested, was justified on the basis of the variability in response of the ores to direct cyanidation, and the fact that the pressure oxidation and subsequent processing steps rendered all environmentally objectionable species innocuous.

Faced with a choice of whether to treat ore or concentrate by pressure oxidation, the major factors to be considered include response of the ore to flotation, the sulphur content of the ore, the content of carbonates and other acid consuming gangue components, and the nature of the autoclaved pulp. In either case, the use of in direct heat exchange for preheating the feed slurry should be avoided since the slurries are essentially gypsum saturated and prone to scaling. Judicious recovery of flash steam and direct heat exchange may enable essentially autogenous oxidation of ores containing ~3% S.

Recovery of acid containing solution from the autoclave discharge pulp in a liquid-solid separation stage prior to CCD washing may meet a portion of the acid requirements before the autoclaving of the ore, and will depend on the relative amounts of sulphur and acid consuming gangue components. Lower sulphur grades and high carbonate levels may justify the added cost of steam and sulphuric acid, depending on the grade of gold and the response of the ore to flotation. Poor response to flotation, very low sulphur grade, and relatively high carbonate levels (e.g. many of the gold ores in Nevada), may warrant consideration of pressure oxidation in an alkaline medium.

In most instances, some degree of upgrading of the sulphides by flotation is preferred. The autoclave size requirements are primarily dictated by the quantity of contained sulphur. Therefore, the processing of a concentrate rather than an ore may not necessarily have a major impact on the autoclave size. It will, however, reduce the size requirements of the other unit operations. The sulphur grade requirements for autogenous pressure oxidation are appreciably lower than those for autogenous roasting. Thus, one of the main objectives of flotation becomes rejection of carbonates and acid consuming gangue to the flotation tailings, which in turn can be utilized in the neutralization circuit.

Pressure oxidation of a lower grade concentrate than required for roasting also offers the possibility of improved gold recovery, by virtue of reduced losses of gold to the flotation tailings, as shown by the results given in Table 1 for an ore upgraded to 8.8 and 17.8% S prior to pressure oxidation. In this case, a high acid consuming capacity of the gangue components, relative to its sulphur content (8.7% CO_2, 2.1% S) made direct pressure oxidation of the ore impractical(14).

Table 1 Effect of Sulphur Grade on Gold Recovery

		% Gold Extraction	
Feed % S	% Sulphur Oxidation	From Autoclave Feed	Overall from Ore
17.8	98.6	96.0	79.8
8.8	99.1	95.6	84.2

The pressure oxidation kinetics of sulphides are strongly dependent on particle size. The results of a series of batch oxidation tests conducted on concentrate ground to a range of particle sizes are given in Table 2(14).

Table 2 Effect of Particle Size on Sulphur Oxidation Kinetics

Concentrate size, K_{80}, µm	110	50	33	20	13
Time for 95% S oxidation, min	92	53	38	30	23

Comparison of the relative capital and operating costs associated with concentrate regrinding versus reduced autoclave size and power requirements indicated an optimum particle size of about K_{80} = 30 µm for this particular concentrate. Although a regrind of an ore will have a similar effect on the oxidation kinetics, the higher costs associated with regrinding of the ore and the possibility of deteriorated liquid-solid separation characteristics may offset some of the benefits derived in the pressure oxidation.

Another factor which may favor the processing of a concentrate rather than an ore is the variability of the ore sulphur content. Although pressure oxidation can accommodate a wide range of feed sulphur grades by means of autoclave feed pulp density adjustment, most efficient operation is achieved with a relatively constant sulphur grade. This is usually more readily achieved with concentrates than with ores.

It is evident, therefore, that the decision of whether to process ore or concentrate and the definition of optimal concentrate sulphur grade and particle size depend on a number of closely interrelated factors which must be thoroughly evaluated.

Pretreatment

Preacidification of the autoclave feed serves to enhance the subsequent oxygen utilization by removing carbonates prior to autoclaving. Another important role of the pretreatment is to ensure sufficient initial levels of acid and iron to promote a rapid initial rate of oxidation to achieve the required temperature profile in the autoclave. After pressure oxidation, there is considerable flexibility in the recovery of acid generated by the oxidation of the sulphides for meeting the requirements of the pretreatment. For most ores, a primary liquid-solid separation (oxidation discharge thickener) will be desirable prior to CCD washing to recover a relatively high strength acid stream for recycle, with higher carbonate ores requiring the addition of make up acid. In the case of concentrates, although the carbonate content is appreciably lower, some degree of preacidification almost always is warranted. The autoclaved pulp may proceed directly to a two stage CCD wash circuit, as shown in Figure 2, and a portion of the diluted acidic waste liquor is recycled to the pretreatment stage.

Solids Recycle for Elemental Sulphur Control

The primary objective of the pressure oxidation is complete oxidation of the sulphidic sulphur to the sulphate form. This is achieved by operating at elevated temperatures, preferably higher than 180°C. Although under these conditions, the sulphur of pyrite is oxidized easily, many sulphides, such as arsenopyrite, pyrrhotite, and base metal sulphides tend to form elemental sulphur as an intermediate or primary oxidation product. Any elemental sulphur that is formed will selectively "wet" the unreacted sulphide minerals, which can then agglomerate.

The high pulp density encountered in the oxidation of ores effectively suspends and disperses any elemental sulphur, such that with adequate retention time the sulphur will be oxidized completely. In the processing of concentrates, at low pulp densities, agglomeration due to the formation of elemental sulphur can lead to severe problems. By recycling a portion of the oxidized solids to build up the autoclave feed pulp density, the sulphide wetted particles can be effectively dispersed and excellent reactivity can be achieved. In addition, the recycle of oxidized solids in effect, provides disproportionate solids retention time(5).

The process flowsheet allows for several choices for the recycle of solids. In the case of high carbonate concentrates, the preferred choice is a portion of the autoclave flash tank discharge slurry, containing a high acidity. Otherwise, a portion of the first wash thickener underflow is recycled. In either case, the slurry is recycled to the pretreatment stage to promote the evolution of carbon dioxide. Prior cooling of the recycle slurry may be advantageous in that it will provide an additional "heat sink" for the heat of the oxidation reaction, allowing for a reduction in the cooling requirements in the autoclave.

Washing of Oxidized Pulp

The pressure oxidation can result in extensive extraction of base metals and appreciable dissolution of gangue components. The primary objective of the wash circuit is the removal of cyanicides and potential slime forming metals such as aluminum, iron and magnesium prior to gold recovery. High dilution of the pulp is employed to promote wash efficiency and flocculant utilization. This allows for the possibility of using a hydrocyclone for washing, and the recovery of coarse solids which may contain unreacted sulphides and possibly refractory gold encapsulated in gangue components. A small regrind circuit for these solids prior to recycling to the autoclave may reduce the regrind requirements for the feed concentrate and enhance the recovery of gold.

Recovery of Silver

Pressure oxidation effectively liberates refractory gold associated with sulphides.

Silver is also liberated initially, but under the prevailing conditions, which favour hydrolysis and precipitation of ferric iron, the silver becomes associated with refractory jarositic species. A recent paper on a study conducted on synthetic ore comprising argentite, pyrite and quartz, suggested that pressure oxidation of the "ore" at 190°C or higher, which favored precipitation of iron as hematite rather than jarosite, led to high recoverability of the silver(17). The studies at Sherritt, conducted in both batch and continuous modes over a temperature range of 170 to 210°C on a wide variety of ores and concentrates have produced iron precipitates ranging from predominantly jarositic to predominantly hematitic. With arseniferous feedstocks, the precipitates also have included ferric arsenate and complex ferric arsenate-sulphate species. Even under conditions favoring predominantly hematite, silver extractabilities have been low, at most attaining 40%, and usually less than 10%. Effective enhancement of the silver recovery can be achieved by atmospheric "conditioning" of the oxidized pulp with lime at 80 to 95°C, to convert sulphate containing iron oxide species to iron hydroxides and gypsum prior to cyanidation(5).

$$Fe(OH)SO_4 + Ca(OH)_2 + 2H_2O \text{ --> } Fe(OH)_3 + CaSO_4 \cdot 2H_2O$$

$$2KFe_3(SO_4)_2(OH)_6 + 3Ca(OH)_2 + 6H_2O \text{ --> } 6Fe(OH)_3 + K_2SO_4 + 3CaSO_4 \cdot 2H_2O$$

It is expected that at least a portion of the ferric arsenate reacts to yield iron hydroxides and calcium arsenate. The effectiveness of the pretreatment has been outlined previously(5) and is further illustrated by recent results obtained in a continuous completely integrated run of 264 h duration, in Table 3(15).

Table 3 Silver Enhancement of Oxidation Residues

	Assay g/t		Extraction %	
	Au	Ag	Au	Ag
Feed to silver enhancement	10.2	71		
Cyanidation of feed	0.58	70	94.2	1.4
Cyanidation after enhancement	0.32	3.4	96.8	95.1

The silver enhancement treatment not only liberated the bulk of the silver, but also resulted in a modest improvement in the gold extraction. Heat requirements for the enhancement are readily met, in that the oxidized slurry can be maintained relatively hot in the CCD wash circuit, and additional heating can be provided by flash steam recovered from the autoclave. Lime requirements are appreciably higher than for a similar pH adjustment on cooled slurry, and the applicability of a silver enhancement will depend on the grades and the relative improvements in silver and gold recoveries.

Environmental

A significant feature of the use of a process employing pressure oxidation for sulphidic feedstocks is that the handling and disposal of environmentally sensitive species is greatly facilitated. Noxious species are converted into relatively stable compounds in the pressure oxidation, neutralization, and silver enhancement operations, and are readily retained in an impoundment area.

In the pressure oxidation, antimony compounds in the sulphidic feed are oxidized, hydrolyzed and essentially completely precipitated. Dissolution of antimony from the oxidized solids in the silver enhancement and leach and CIP operations is virtually negligible (<0.1 mg/L). Arsenic is oxidized sequentially to the soluble arsenite and arsenate states and is subsequently precipitated as ferric arsenate or complex ferric arsenate sulphate species. The extent of arsenic precipitation is dependent on a number of factors, including arsenic level in the feed, the relative amounts of iron and arsenic, solution acidity, and operating temperature, but generally between 85 to 95% of the arsenic is precipitated. Dissolution of the arsenic from the oxidized solids in subsequent processing steps is virtually negligible. In the CIP or CIL circuit, entrainment or entrapment of oxidized solids will result in trace levels of arsenic on the carbon. Treatment of the acidic water from the CCD wash in the neutralization circuit usually removes arsenic to <0.1 mg/L. The low levels of arsenic, compared to solubility levels of 2 to 3 mg/L As often quoted, and 75 mg/L sometimes quoted(18) in the literature for ferric or calcium arsenates are attributed to a high Fe:As ratio(19) in the feed liquor and the large mass of precipitated hydroxide slimes which effectively absorbs or chemisorbs the arsenic. Recently reported commercial experience on the handling of related arsenic containing wastes has confirmed their suitability for impoundment(20).

Base metals including cadmium, cobalt, copper, nickel and zinc are largely extracted during pressure oxidation and are removed from the oxidized solids in the wash circuit. Consequently, these metals are present in only trace amounts in the gold recovery circuit, resulting in lower cyanide consumption, cleaner leach solutions and lower contaminant levels on the carbon. The metals in the acid stream are precipitated to trace levels as their respective hydroxides or hydrated oxides in the neutralization circuit. Lead sulphides are converted to the relatively insoluble sulphate and jarosite species during the pressure oxidation. Dissolution of the lead from the oxidized solids in a silver enhancement treatment and in the leach and CIP solutions is negligible.

Gold ores may often contain appreciable levels of mercury. During pressure oxidation, it is expected that the mercury containing species are largely converted to mercuric sulphate which, under the prevailing conditions, is precipitated as a mercuric jarosite or is otherwise incorporated in the complex iron containing species. Extraction of the mercury by cyanidation of the oxidized solids may be variable, but is lower than that obtained by cyanidation of unoxidized feeds. The silver enhancement treatment also enhances the cyanide extraction of the mercury, but not to the same extent as that of the silver. Thus, although the pressure oxidation renders the bulk of the mercury refractory, provisions are required for control of mercury during the recovery and refining of the gold and silver.

Gold ores often contain selenides and tellurides. Although the feed stocks studied to date at Sherritt have contained only minor levels of these species, the results indicate that in the pressure oxidation, selenides and tellurides are extensively oxidized and are essentially precipitated from solution with the iron compounds. Dissolution of selenium and tellurium from the oxidized solids in subsequent processing steps is negligible.

The autoclaving process oxidizes essentially all of the sulphur to the sulphate form. A portion reports to the solids, as alunite, jarosites, basic ferric sulphates and calcium sulphates, while the remainder is present as acid and associated with soluble metals in the acid liquor. Treatment of the acidic solution in the neutralization circuit precipitates the sulphate as gypsum. The more labile sulphate containing species in the oxidized solids are also readily converted to gypsum during the pH adjustment prior to cyanidation; a silver enhancement treatment results in conversion of the bulk of any jarositic sulphate to gypsum. Thus, ultimately the sulphur in the feed stocks is discharged either as gypsum or a jarosite-gypsum mixture.

After precious metals recovery from the oxidized solids, the barren pulp will usually be dewatered, to recover cyanide and reduce the cyanide content in the barrens. It can then be combined with dewatered sludge from the neutralization circuit for deposition in a tailings impoundment area. If it proves necessary to detoxify the cyanide containing barren slurry prior to impoundment, effective removal of cyanide can be achieved by treatment with the metals containing acid liquor from the wash circuit. This concept was demonstrated in the most recent integrated studies on the Porgera concentrates, where the barren CIP slurry was added directly to the initial, acidic stages of the neutralization circuit(15).

Recovery of By-Products

The earliest work involving pressure oxidation of refractory gold containing sulphides stems from work on cobalt arsenides and sulphides. Later work on pyrite containing uranium and gold ores showed a modest enhancement of gold recovery after pressure oxidation(1). In both cases, the enhanced gold recovery was considered a side benefit. Although the recent work on the pressure oxidation of refractory sulphidic feedstocks has been primarily concerned with gold and silver, the possibility of recovering associated metals warrants consideration. Pressure oxidation of uranium containing gold concentrates will not only effect high extraction of the uranium, but will also provide acid and iron containing solution which can be utilized for leaching uranium ore. Base metals, which are effectively extracted in the pressure oxidation, may be recovered at an intermediate stage of the neutralization circuit. This offers the possibility of treatment of otherwise unattractive feeds such as gold or silver containing flotation middlings or low grade concentrates recoverable from tailings of flotation operations on cobalt, copper, nickel and zinc operations, in conjunction with the major gold feeds. Although a process employing acid pressure oxidation ultimately converts arsenic to ferric and calcium arsenates for impoundment, the presence of a suitable market for arsenic could make its recovery feasible. Work is currently under way at Sherritt on this and other potential applications.

CONCLUSION

After testing over 80 ores and concentrates and designing "tailor-made" processes for ores, concentrates and "upgraded" ores, Sherritt has concluded that the use of pressure oxidation is often the method of choice. Dominant factors affecting the economics are power costs, reagent costs, capital costs and ore grade and composition. Given a reasonable minimum ore reserve and a gold price of about US $11 000/kg (US ~$340/oz), then a very economically attractive project should be realized if the sulphur to be oxidized:-refractory gold ratio is equivalent to 10 000:1 (e.g. 3% S for 3 g/t Au). Under favourable circumstances, much higher ratios can also give positive results.

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